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Editor

CONTENTS

Skelly Modernizes Special Products Plant 3
J. C. Albright

Comparison of Costs of Operating Shell Stills and Pipe Stills .. 7
W. L. Nelson

Design of Continuous Treating Plants 11
A. W. Trusty

The Refinery Transportation System 15
L. C. Trescott

Selecting and Maintaining Electrical Equipment 23
V. R. Chadbourne

A Simplified Method for Absorber Design 30
G. T. Atkins and W. B. Franklin

Separation of Solvents and Water in Refining and Dewaxing Lubricating Oils 33
Eugene R. Smoley and Wheaton W. Kraft

Coking Operation, 168 Days on Stream 38
J. C. Albright

Departments

The Look Box ..... 1 and 37a

Plant Activities 42a

Business Notes 45a

Ideas 48a

Science and Technology (Abstracts) 50a

New Equipment for the Modern Plant 58a

Catalogs and Bulletins 61a

Index to Refiner, Vol. 14, 1935 ..... 63a

Advertisers' Index 72a

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VOLUME 15, No. 1 JANUARY, 1936

REFINER
AND NATURAL GASOLINE MANUFACTURER

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The Look Box

Engineering Economics

IN this issue begins presentation of a group of articles or discussions which are classified under the general heading of Engineering Economics. The articles presented at this time, and those to be published in future months, are the outgrowth of a thought expressed in these columns last July. That thought, briefly, was that the engineering economist who applies his engineering knowledge to the economics of unit operation is that much ahead of the engineer or the economist, for he is the "fellow who figures things out and makes them work —at a profit."

A large group of well known men in the industry have contributed or are planning to contribute to this group of articles, from their knowledge and experience, engineering and technical information which will prove of immense value to those in the industry interested in its rapid technical and engineering development.

Nor is the topic restricted to those who are now cooperating in the preparation of these important discussions. The Refiner cordially invites further contributions and cooperation from those who feel that they have something to say that is of value to their fellow members of the industry.

Examples of the class of discussions which are presented in the Engineering Economics series found in this issue include, W. L. Nelsons' discussion of shell stills versus pipe stills; V. R. Chadbourne's important analysis of selection, installation and maintenance of electrical equipment; L. C. Trescott's interestingly prepared paper on the refinery transportation system, and A. W. Trusty's timely article on design of continuous treating systems.

Future articles will deal with virtually every phase of refinery and natural gasoline plant design, construction, operation and maintenance, and the scope of the group will include virtually all departments of the plant from the loading rack to the front gate.

Coming, for example, are articles dealing with cracking in its various phases, polymerization, solvent processes and processing, experimental methods, gas gathering systems and gas handling, compressors and compressions, modernization of gasoline plants, design of proper storage facilities, percolation filtration, solvent recovery, use of exhaust steam in distillation, heat exchange, condensation, and so on. The scope is ever widening as more engineers and technologists become interested in the program and offer suggestions and contributions. Almost all of them are good, all are appreciated.

The technologist, chemist, scientist, or research worker, whatever may be his title, is of the type that develops new processes, new adaptations or applications of old processes or combinations of new and old. The engineer provides the all-important engineering knowledge that leads development through experimental to commercial application. Both must work together if gains are to be made. And further, both must know and apply the economics of unit operations if processes are to be pieced together into a smooth running assembly that will make more money because of more efficient performance. It is on the common ground of economics, after all, where those who are developing and progressing must meet and go forward.

The scientist, technologist, chemist; the engineers—chemical, mechanical, civil, power, electrical, designing, and others—these have taken strong hold on the industry and during the past 10 years propelled it further along scientific lines than was dreamed of in, say, 1915. In the profusion and often confusion of rapid development, that which has proven most successful has seen a deal of attention from the standpoint of "pennies" in cost of operation. Which brings us back to economics, and to Engineering Economics.

Outlook

FOLLOWING closely the indications that appeared during the closing months of 1935, the petroleum industry, it is felt, is entering the new year with much to indicate a sound condition generally, and with a distinctly promising outlook.

Predictions coming from various authorities have been made to the effect that 1936 will see a new all-time record set in consumption of oil, and with total demand expected to be possibly six percent greater than the demand during 1935. Granting correctness of these estimates, the petroleum refining industry has made, and is making no error in the marked activity of new construction and equal or greater activity in modernizing and rehabilitating its processing equipment. And this work, as is well known, is going on in practically every department and phase of manufacturing. Likewise, the natural gasoline manufacturing industry is active in the revamping and modernization of its processing facilities, as well as the establishment of new installations.

Meanwhile, inventories are the lowest for nearly a decade. Current production of crude petroleum conforms closely with consumption. This has brought about a condition where markets are comparatively strong, with prices generally at favorable levels. Many authorities state, as this is being written, that increases in both crude petroleum and refined products prices are apparently in store.

Oil company earnings improved further during the past year, having been the best since 1930; and such earnings are expected to show additional expansion during 1936, as evidenced by the popularity of oil company securities lately.

Optimism over the oil industry's outlook is felt by leaders within the industry, as well as by investors and others. Heads of some of the larger companies have expressed this confidence in recent interviews. Others have expressed it by adopting more liberal policies in paying dividends to stockholders, paying bonuses to employees and in some instances increasing salaries.

The Role of Nickel

TIME was when oil refinery specifications included much mention of "Black Merchant Pipe." So much in fact, that in 1918 a design engineer remarked that he would like to meet the black merchant, since he had seen quite a lot of mention of his pipe. Progress brought high temperatures, high pressures, and nature brought more corrosive oils while the industry learned more of corrosion and how to combat it. The pipe of the olden days marched along with demand, changed complexion, physical and chemical properties and today the refining of petroleum is carried out through the means of a wide and important range of metals—steels into which are alloyed nickel, chromium, molybdenum, silicon in varying percentages. Of course, growing knowledge of metals has brought more usage and more proper usage of cast iron, alloyed cast irons, wrought irons, plain carbon steel, and aluminum and its alloys.

That the role of nickel in refining would become of such importance as reached at the present stage of development, was little dreamed of 20 years ago. No one would have agreed in the early days of continuous shell-still distillation systems that nickel would be consumed in alloy with steel, chromium, molybdenum, in tonnages reached to this time—and the end is not yet.

At sub-zero temperature ranges at which some processes for dewaxing of lubricating oils are carried out, many metals lose much of their impact strength. Steels containing two to five percent nickel retain a good part of their room temperature impact resistance in this range—so there follows a widespread adoption of nickel steels with low carbon content for large vessels used in these processes. [*Continued on page 35a*]

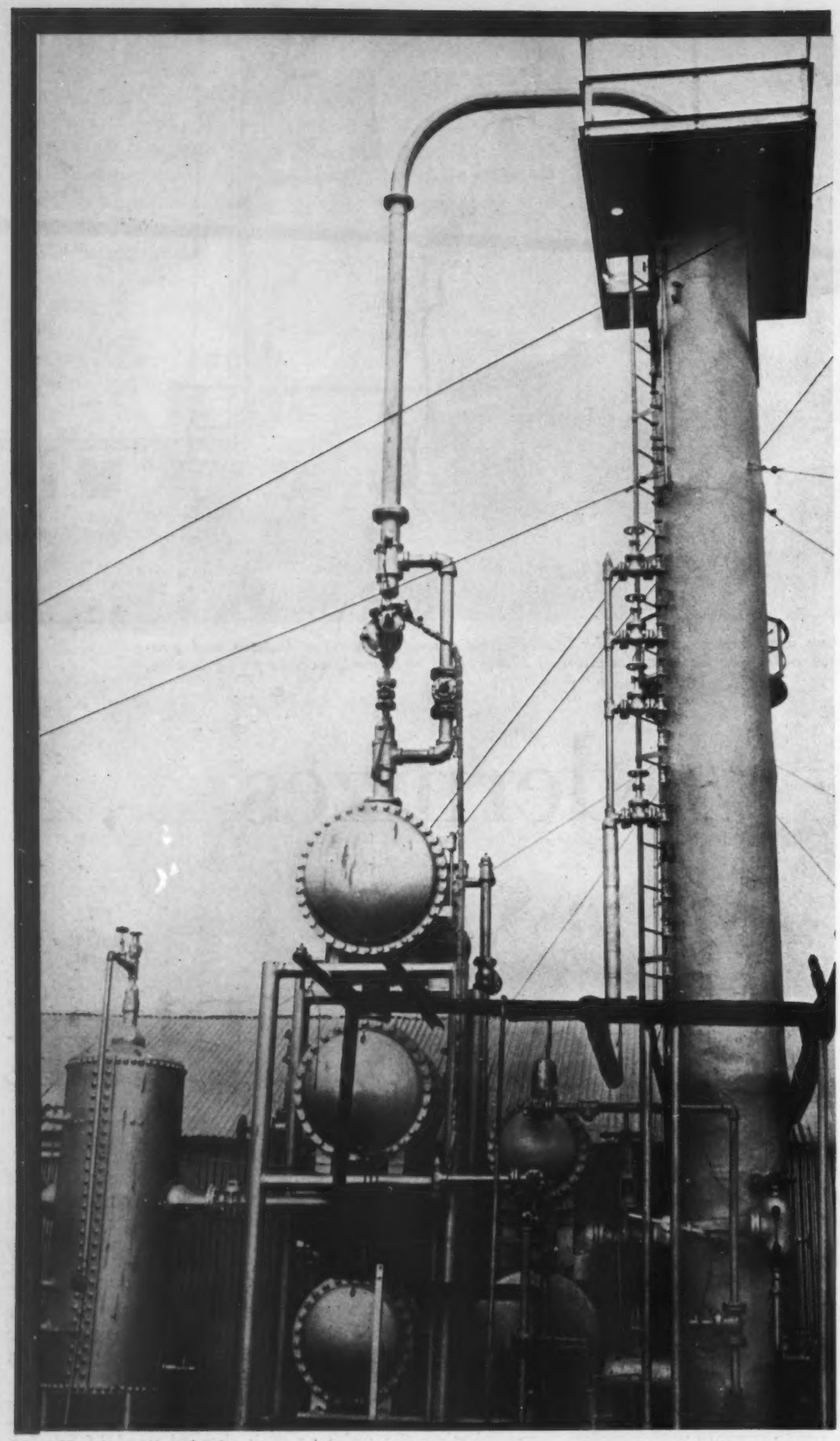
Three fractionators used in Lyman Plant by Skelly Oil Company. Two for butane and propane removal, and other for hot fractionation of compressed rich gas in recovery of gasoline.

Skelly Modernizes *Lyman Special* Products Plant

GASOLINE plants, thought to be the last word in efficiency a few years back, now are being modernized so the cost per gallon can be reduced, and at the same time, extract more gasoline from a given amount of wet natural gas. The industry generally, is following this practice by the addition of more modern absorption units, larger and more efficient heat exchangers, fractionators which will cut the product closer to a given point, and changing from compression to absorption, a combination of the two, or from absorption to gas fractionation operation. Some have found that it is much more inexpensive in the long run to completely revamp a plant than it is to try to operate with obsolete equipment, badly in need of extensive repairs.

One particular instance of this kind is that of Skelly Oil Company at its plant located at Lyman, in the north end of the Burbank field, Osage County, Oklahoma. This plant was built several years ago, using the absorption system then generally widely used. The system of erection, layout and general operation was about the same as that of most others in the industry, and the plant consisted of field lines, compressors, using about 35 pounds discharge pressure, absorbers and run tanks. Gasoline was weathered to the required specifications in storage tanks, but a short time later, fractionators were installed which led to more efficient separation of the undesirable propane and butane, and left a large amount of marketable gasoline.

The major operations of recovering the gasoline contained in the gas were followed for several years, reconditioning the units as they showed need of repairs, and adding fractionator columns for special products. During the most of 1934, and a part of 1935, the entire absorption, or gasoline recovery part of the plant began to show signs of extensive wear, and it was decided that it would be more economical to completely rebuild



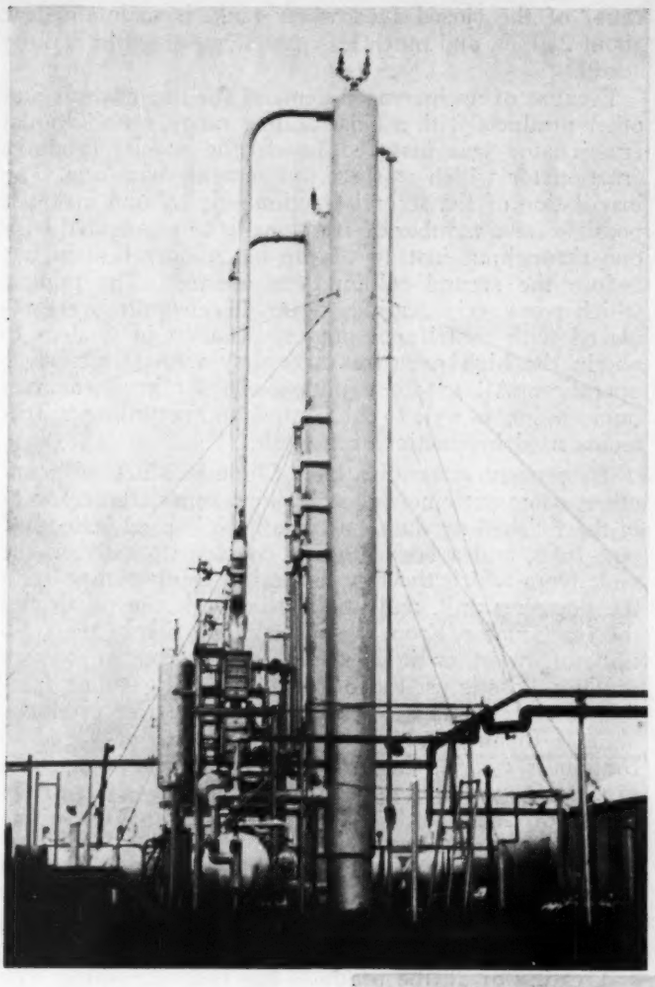
Close-up of reflux condensers installed when the plant was changed from absorption to hot compression operation.

the plant than it would be to operate it by continually repairing the various parts.

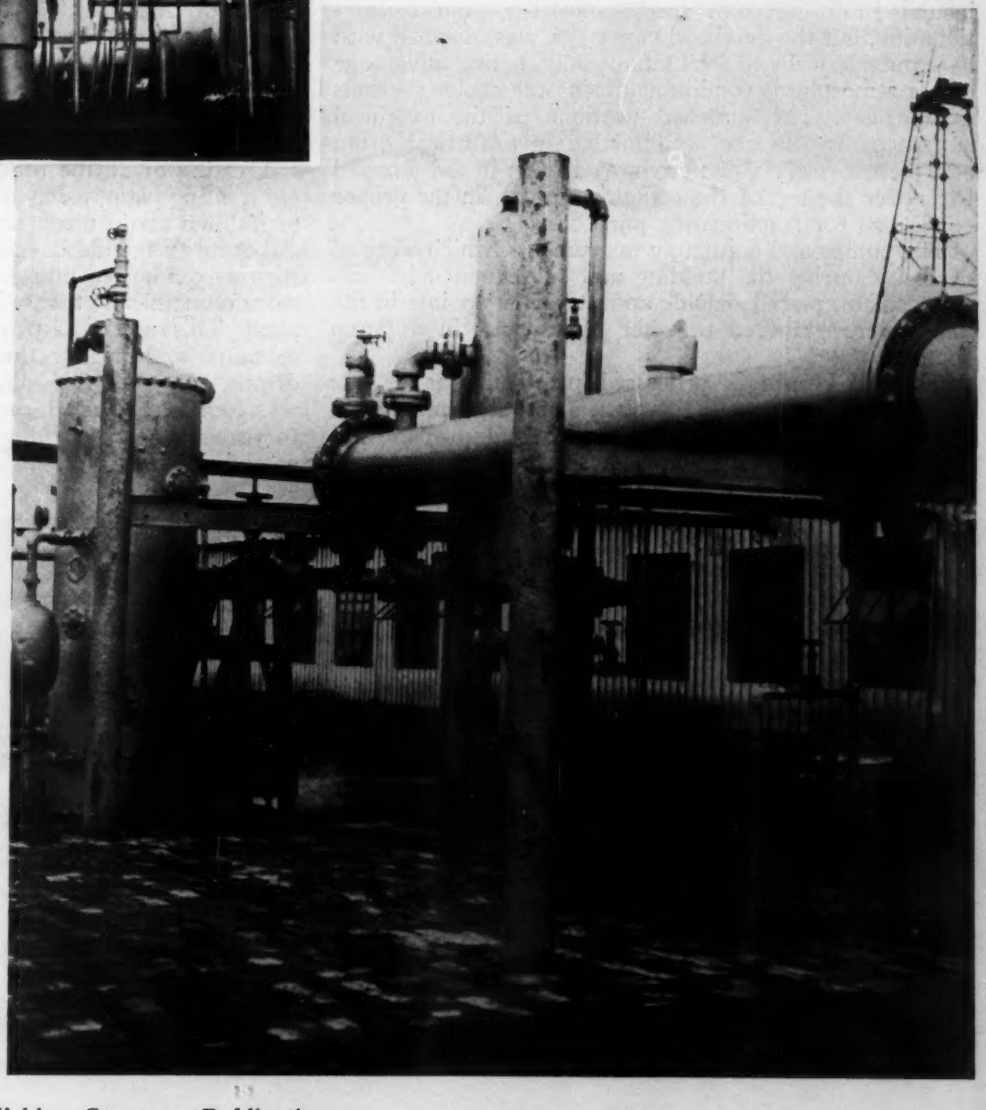
As a result the absorption part of the plant was discarded and the compressors refitted with smaller cylinders, and the pressures boosted to 175 pounds, instead of the ordinary 35 pounds used when operating the absorption plant. The plant site was cleared of all the cooling towers, absorbers, stills, and graded for the newer equipment which was installed as fast as possible in order that the shut down time would be shortened. The result is a modern plant which operates much more cheaply than the old one.

The engine room was left in practically the same condition as formerly, with the exception of reconditioning both the power and compression cylinders, and removing one of the low-pressure cylinders for each twin engine and substituting a smaller bore for the required higher pressure operation. The old louvre tower, which had seen service in cooling the gas and mineral seal oil during the absorption life of the plant, was torn down. As the newer type of gas fractionation operation required no particular cooling of the gas prior to its entry into the fractionation column, the space occupied by the louvre tower and two-inch pipe coils was used for other units.

The primary fractionating column used during the last few years when the plant was operated as an absorption unit, and which was used to separate the butanes and



Special towers for manufacture of close boiling range products.



Special built propane refrigerating unit.

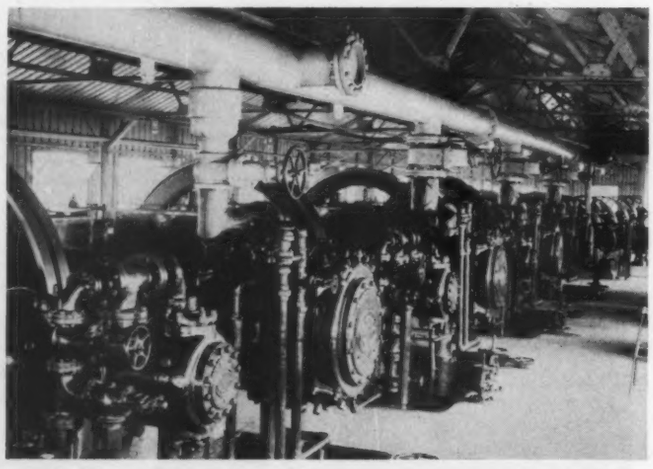
the propane from the bulk of the "make," was reconditioned, and fitted with the necessary reflux coolers, exchangers, and the like to make it operable as a column in hot gas fractionation. The gas, direct from the high pressure compressors, is introduced into this column at the point where it is most practicable, and the column refluxed with condensate to complete the fractionation of the desirable hydrocarbons from the undesirable ones, particularly ethane and methane.

The column is operated at the desired pressure and top temperature so the maximum amount of butanes and propanes can be extracted as well as the lower boiling fractions, because, in the operation of this plant as a whole, the propanes and butanes are desirable fractions, later to be liquefied and purified for various special products. One special feature found in this column is the location of the valve which controls the column pressure. Instead of being placed on the accumulator drum, it was placed in the overhead vapor line between the top outlet of the column and the first condenser section. This makes possible a steady, nonfluctuating pressure maintained at all times on the column, independent of the temperature of the condensers or the amount of condensate trapped in the accumulator drum. A by-pass was installed on the vapor line which is in operation at all times, but insufficient for complete operation, and thereby forces the regulating valve to take care of the excess amount of gas coming from the top of the column.

The column, as well as the reboiler at the base was insulated to preserve the heat applied for complete fractionation, but the overhead vapor line was installed with bare outside walls to facilitate cooling, taking advantage of the atmospheric conditions during the cooler portions of the day. The liquefied portions of the overhead stream of vapors are accumulated in a vertical drum and the necessary volume removed from it and pumped back over the top of the column to maintain the proper conditions for fractionating purposes.

The pumps and regulating instruments, which were of necessity during the building and subsequent additions to the plant, placed outside and at various points in the yard, were gathered together and placed in a pump room.

The raw make is processed practically as it was during the operation of the plant as an absorption unit, and the special fractions of butane and propane are handled as they had been. The boiler room was overhauled as well, and the open top hot well, or elevated tank, which was installed for conditioning the boiler feed water, was removed and in its place a pressure vessel was installed into which the raw water is pumped, and into which all condensed steam is collected. The vessel is equipped with a liquid level instrument, with the float located so that when the water in the vessel is not maintained at the proper level a stream of make-up water flows into it. The temperature, because of the closed feed water tank, is maintained at about 220°F., and much less gas is consumed in making steam.



Gas engine driven compressors redesigned by the addition of high pressure gas cylinders when Lymon plant was rebuilt.

Because of an increased demand for the solvents and other products with a close boiling range, an additional fractionator was installed beside the special products fractionator which was erected several years ago. The installation of the second fractionating column makes it possible for a number of fractions to be segregated with one throughput instead of the batch operation in use before the second column was erected. The pumps, which were reciprocating, steam driven units, were replaced with centrifugal pumps, installed in tandem to obtain the high pressures necessary with this type of operation. All of these pumps are set in a common pump room, as well as the control and recording instruments used in fractionating work.

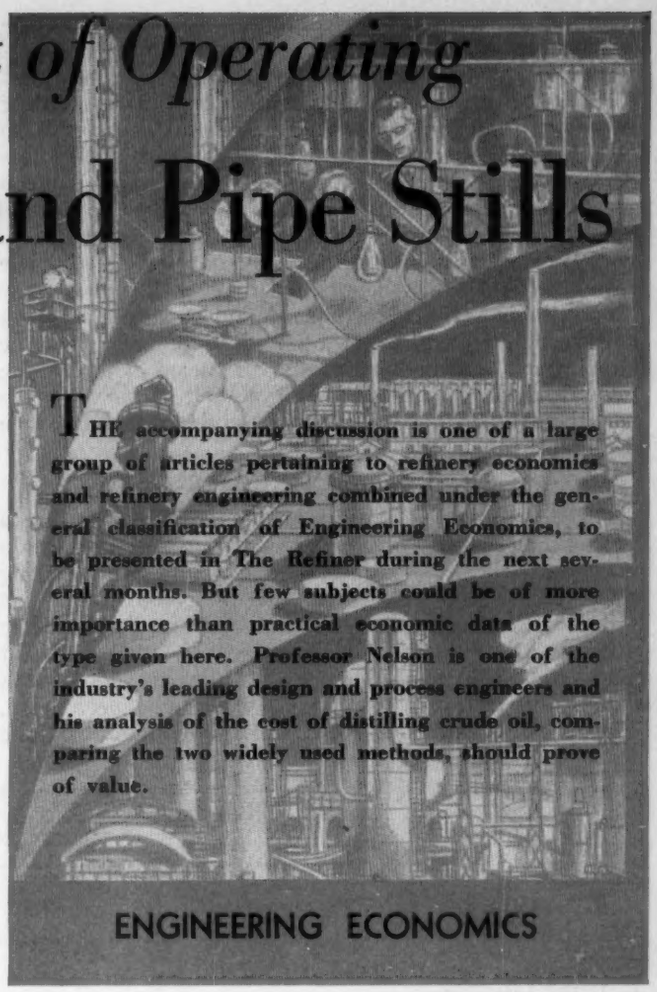
To prevent contamination of one product with another, when extremely close boiling ranges are necessary in the finished product, gasoline hose is used instead of pipe lines, and a fresh hose is coupled up between the tank from which the raw natural is being pumped and the pumping unit, and another between the pump and the fractionator. From the outlet connection of the fractionator, whether it is overhead fraction or bottom product, hose is used also to transfer the gasoline fraction to storage. This product, or rather, these products, because of the multiplicity of them, are packaged in containers ranging in size from a tank car to a quart can. Drums are used extensively in shipping some of the special products, such as rubber solvent, etc. These two fractionating columns are used for the manufacture of "tailor made" products, made from the debutanized natural gasoline to order, some of which require a certain boiling range, with a designated initial and end point.

Treating of all the products has been simplified. The old treating plant, using some form of hypochlorite, was torn down and a much smaller unit constructed, not so elaborate as the old one, because it has been found that treating can be accomplished with a simplified unit much more economically than with a larger, more cumbersome plant. This unit was placed near the special products columns, and is new throughout—fittings, piping and drums. As with the balance of the recovery plant, it was found to be much more economical to rebuild than to recondition often, due to worn pumps, eroded and corroded piping and leaking valves and fittings.

A program is under way to further fractionate the product, and an additional three columns are being erected which will place the company in a position to make any thing from liquefied propane to the heavier fractions, relatively pure, or blended after being fractionated from the raw make. Gasoline made in South Burbank, as well as the recently acquired Carter "9" plant will be processed in the Lyman plant together with the production regularly made there.

Comparison of Cost of Operating Shell Stills and Pipe Stills

W. L. NELSON, ***Consultant***
Professor of Petroleum Refining,
University of Tulsa



THE accompanying discussion is one of a large group of articles pertaining to refinery economics and refinery engineering combined under the general classification of Engineering Economics, to be presented in The Refiner during the next several months. But few subjects could be of more importance than practical economic data of the type given here. Professor Nelson is one of the industry's leading design and process engineers and his analysis of the cost of distilling crude oil, comparing the two widely used methods, should prove of value.

ALTHOUGH shell still operation is conceded to be a costly method of processing, many refiners continue to use shell stills for topping crude oil and for other distillation work. Probably more than half of the refineries operating in 1935 are using shell stills to process a part of their crude, although it has been estimated that only about 33 percent of all petroleum distillation now is processed in shell stills. It should also be remembered that much of the percentage consists of the redistillation of pressure distillate, naphtha, pressed distillate, and bright-stock solution.

The purpose of this paper is to compare the costs of pipe still and shell still operation and to determine the reasons for the difference in cost. There must be plausible reasons for the fact that so much shell still capacity is still in use and in part these reasons will be suggested. Perhaps the most powerful influence for the continued use of shell stills is the lack of funds of many refiners or their lack of confidence in the immediate economic condition of the industry. Many of these refiners are convinced that their operations are expensive but they cannot feel encouraged to incur additional expenses while market conditions are unsatisfactory. This is a particularly difficult situation to be in, since they must compete with refiners with modernized plants and the competition becomes more keen each year because of new pipe still installations. However, the cost analysis which follows indicates that the cost of operating a well kept shell still battery is not prohibitive. In other words, the loss by the operation of a shell still battery depends entirely on the condition of the shell still equipment and particularly the effectiveness of the fractionating system. Further, those refiners who have not written off their investment in shell stills will have a big obsolesence charge to carry. Shell stills must be discarded before the next business depression because new pipe still competition is appearing each year.

By means of this paper, shell still operators should be aided in analyzing their particular operation, to determine wherein their costs are too great.

OPERATING FACILITIES

The discussion which follows deals particularly with the operating features of the two equipments, and what these operating facilities mean in dollars and cents will be reserved for a later part of the paper. Many of the advantages of pipe still operation are of an intangible nature and cannot be evaluated in dollars and cents except as they affect the general morale of the plant force, cost of supervision, and savings on repairs. We refer particularly to fire hazards, the safety of plant workmen, and the advantages of a continuous controllable operation.

The major advantages of a pipe still over a shell still are:

(1) *The pipe still is best suited for continuous operation.* This permits a rigid control of the operating conditions such as rate, maximum temperature, amount of steam used, and firing conditions. Continuous operation also permits the use of extensive heat exchange and an increased thermal efficiency.

(2) *The pipe still is particularly adapted to the practice of flash-vaporization.* By flash-vaporization it is possible to distill a much larger amount of the crude oil without causing decomposition or the use of an excessive amount of process steam. Thus, most crude oils can be topped at a maximum temperature of 690°F., which requires no steam, if a pipe still is used, whereas a temperature of about 800°F. is necessary in a shell still battery or a large amount of steam to reduce the final temperature to the customary 630°F.

(3) *Less decomposition occurs in a pipe still.* Although the rate of heat transmission is higher in a pipe still than in a shell still, the heating can be controlled by means of the velocity, tube arrangement, and flame conditions, so that localized hot spots and decomposition do not occur to a large extent. This results in greater

yields of the valuable lubricating oil products and a better separation between the lighter products.

(4) *The rate of heat transfer is higher in a pipe still.* Although the heating surface in a pipe still is more expensive than in a shell still, the rate of transfer is so much greater that the total cost of the pipe still is less than the cost of shell stills. The advantage is even greater in the entire distillation unit because more fractionating towers are required with the shell stills.

(5) *Pipe stills reduce the fire hazard.* The large volume of oil that is contained in a shell still battery constitutes a serious fire hazard and a danger to plant workmen. Pipe stills are built so that a fire can be quickly extinguished by means of steam whereas the several hundred barrels of oil in a shell still may escape into the refinery yards and drains and seriously menace the entire plant.

PROCESSING COSTS

Although overall operating costs taken from hand books, articles, etc., are of questionable value, they do serve to indicate the general magnitude of operating costs. These data are called questionable because the source of the information is not entirely clear, widely different stocks may be discussed by the different authorities, the items that are included are not always listed, and different equipments may have been used. The Cross Handbook[1] gives the following indications for a 2000-barrel plant:

	Dollars per Bbl. of Crude Oil	
	Pipe Still	Shell Still
Distillation	0.10	0.30
Cracking	0.27	0.35
Coking	....	0.50

[1] Handbook of Petroleum Asphalt and Nat. Gas No. 25, page 344, K. C. Testing Laboratory.

Other authorities give costs for pipe still distillation between seven and 11 cents per barrel of crude oil but few of them venture to discuss shell still costs.

The items that contribute to this difference in cost may be summarized as follows:

(1) *Yield.* Large losses are not involved unless the fractionation system is inadequate. However, the tower systems on most batteries are inadequate.

(2) *Steam.* Much more steam is required for a shell still battery because of the non-flash conditions that exist in a shell still.

(3) *Fuel.* The thermal efficiency of the pipe still is much greater, and further, because of continuous operation, heat exchangers are more effective if used with pipe still equipment.

(4) *Repairs.* The repair of old shell still equipment is an important item.

(5) *Operating Labor.* In large equipment, the cost of operation of pipe still equipment is much smaller.

(6) *Capital Investment.* New pipe still equipment is much cheaper than new shell still equipment. However, most shell still equipment has been depreciated to a large extent and hence the depreciation charges are usually about the same on both equipments.

(7) *Difficulty in Treating.* With inadequate fractionating towers the distillates often require additional acid.

These items will be discussed in detail and the costs for a 3000-barrel crude oil topping plant will be computed. A typical True Boiling Point analysis of a Mid-Continent crude oil is shown in Figure 1. In addition to the True Boiling Point distillation curve (No. 1), the Mid-percent gravity curve (No. 2), and the Flash Vaporization curve (No. 3), a special curve of the still temperatures in the several shell stills is shown in Curve No. 4. About 13 shell stills will be required and the operating conditions in these stills are shown in Table 1.

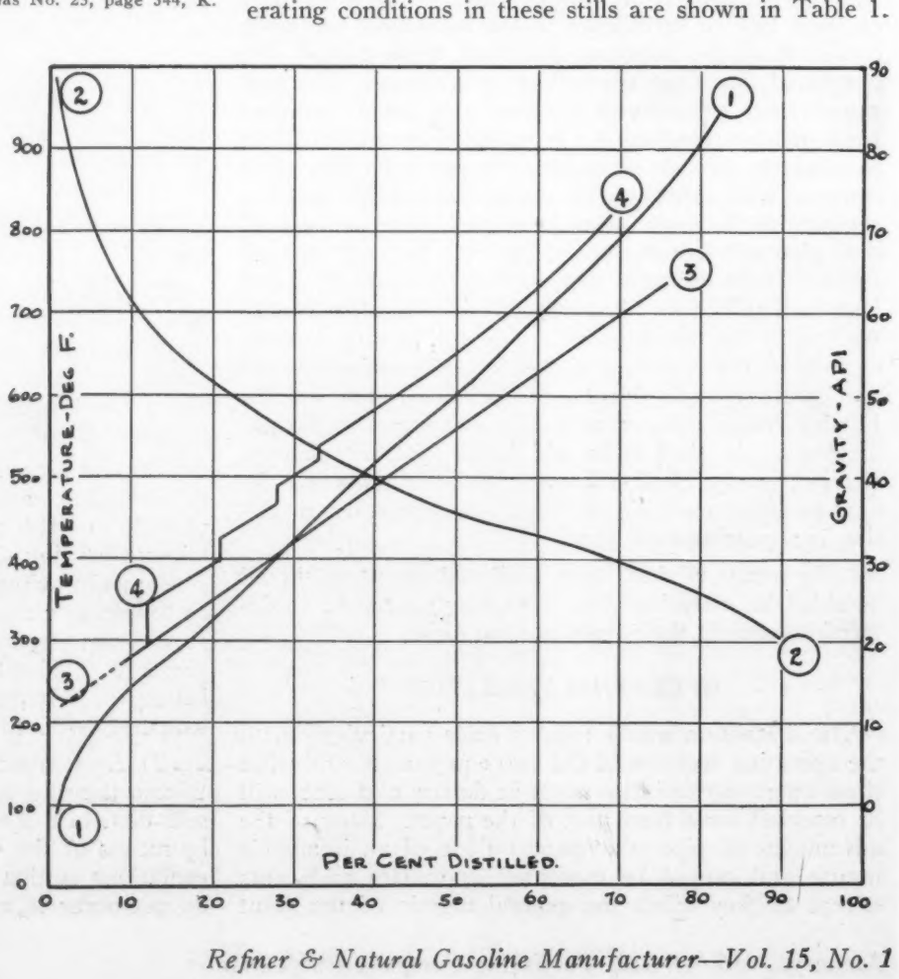


FIGURE 1

37° A. P. I. Mid-Continent Crude Oil

TABLE 1
Operating Conditions in Shell Still Battery

Still No.	Material Distilled	Actual Still Temp. °F.	Still Temp. If no Steam Used °F.	Amount of Steam Lb. per Hour	Notes
1	Naphtha.....	290	290	....	
2	Naphtha.....	396	396	....	
3	Naphtha.....	467	467	....	
4	Naphtha & kero...	520	520	....	high reflux
5	Kerosene.....	570	570	....	high reflux
6	Gas oil.....	605	605	....	
7	Gas oil.....	610	642	187	first steam
8	Gas oil.....	630	690	235	
9	Gas oil.....	630	723	286	
10	Heavy gas oil*....	630	746	284	
11	Heavy gas oil*....	630	774	440	
12	Heavy gas oil*....	630	800	590	
13	Heavy gas oil*....	610	820	1010	high reflux

*In a lubricating oil plant, this would constitute the lightest part of the wax distillate.

The operating conditions of the pipe still will not be given in detail because they are unnecessary. The vaporizer temperature will be about 665°F., and the pipe still outlet 700°F. No steam is required to reduce the temperature at the vaporizer but steam is used to strip the kerosene, gas oil, and reduced crude oil.

The yields by the two methods of processing are indicated in Table 2.

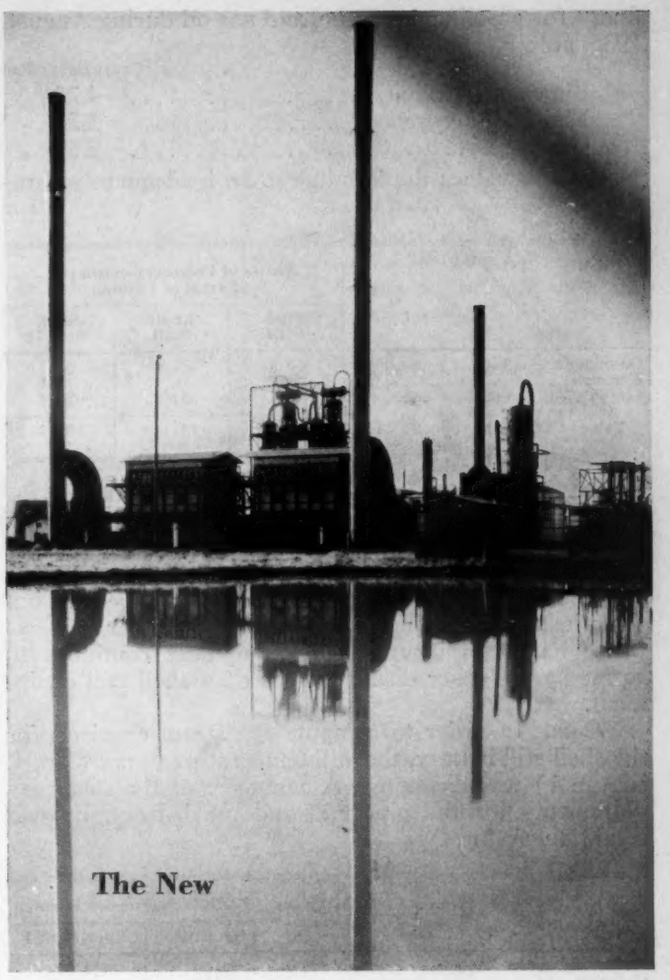
TABLE 2
Yields When Processing 36° A.P.I. M. C. Crude Oil

Material	Properties	By Perfect Separation Percent	Pipe Still Percent	SHELL STILLS	
				I Normal Condition Percent	II Inadequate Towers Percent
Naphtha..	410 E.P.	31	30*	29.0	27
Kerosene..	42 °A.P.I.	8	10*	10.5	13
Gas Oil...	43.5 °A.P.I.	30	29	29.5	29

*Perfect fractionation not desired. Must produce burning kerosene.

Several sets of actual plant runs were used in preparing Table 2. In some cases the separation was even poorer than shown for shell stills with adequate towers.

Yields. As indicated in Table 2, the loss due to an imperfect separation of the products depends to a large extent on the particular equipment. The separation that is attained is greatly dependent on the fractionating system and as a rule the tower systems on shell still batteries are inadequate. The market prices at the re-



The New

The Old

finery for gasoline, kerosene, and gas oil during August, 1935, are:

	Cents Per Gallon
Finished gasoline	4.75
Finished kerosene	3.5
Gas oil	2.5

At these values the loss due to an inadequate separation is:

	Value of Products-cents per Barrel of Crude		
	Pipe Still	Shell Still	Shell Still II
Gasoline	59.9	57.9	53.8
Kerosene	14.7	15.4	19.1
Gas Oil	30.4	31.0	30.4
Value of Distilled Products	105.0	104.3	103.3

Thus, the loss due to an inadequate separation in the shell still battery ranges from 0.7 cents per barrel to 1.7 cents and even more, depending on the equipment.

Towers can be provided for the shell still battery that will produce nearly the same yields as a pipe still but their installation is not justified because after this expense has been incurred the equipment continues to suffer by the other disadvantages of a shell still equipment.

Steam. In order to compute the steam erquired for the shell still battery, the still temperature curve (No. 4, Figure 1) was computed. A summary of the steam requirements for the pipe still and the battery in good condition, follows:

Item	Pounds Steam per Hour	
	Pipe Still	Shell Still I
Process	....	3,032
Stripping	1,127	490
Burners	276	620
Pumps	700	770
Legitimate Losses	210	740
	2,313	5.652

At a steam cost of 30 cents per million B.t.u. the charges amount to 0.6 cents per barrel for the pipe still and 1.47 cents per barrel for shell still I.

Fuel. The thermal efficiency of the best non-recirculation pipe stills seldom exceeds 55 percent and this was the figure used for Shell Still I. The pipe still efficiency was used as 70 percent although many stills are operating at a higher efficiency. Furthermore, the temperature attained by heat exchange, on a shell still battery, seldom exceeds 280°F., whereas the heat exchange system on a modern pipe still usually raises the crude oil temperature to 340°F. The value of refinery fuel was considered to be 40 cents a barrel.

At these conditions the fuel for the 3000-barrel pipe still unit amounted to only 0.88 cents per barrel whereas the shell still I fuel expense was 1.2 cents per barrel.

Operating Labor. A pipe still unit properly equipped with control instruments requires only one operator per shift and one helper. The same applies to a shell still battery in good condition but as a rule three men are required. Operating labor was valued at an average of 45 cents per hour. For these conditions the direct operating charges are 0.72 cents per barrel for the pipe still and 1.08 cents per barrel for shell still battery I.

Repairs. This is an item that must be determined by each refiner. Nevertheless, shell still batteries are usually old and constant repairs are necessary. Among the repair items for a pipe still are—unskilled labor, new tubes, gaskets for tubes and tower manholes at each shutdown, removal of dangerous or broken fittings, tube rollers, power for compressed air, and steam.

For a shell still—unskilled labor, new still bottoms, welding of these bottoms, gaskets for towers and stills, brickwork repair, cleaning of still bottoms, repair of valves and fittings, repair of insulation, cleaning condensers, and steam. For average conditions these items amount to about 0.502 cents per barrel for the pipe still and about 1.56 cents per barrel for shell still I.

Capital Investment. This also is an item that depends entirely on the plant management and accounting. The cost of new pipe still equipment today is about $120,000, or $40 per barrel. The cost of a new shell still battery of 3000 barrels capacity would be far higher but as a rule, refiners do not list their obsolete equipment at high figures. If the existing shell still equipment is valued at $30,000 and it is to be depreciated in four years the charge per barrel is about 0.71 cents. The investment charge for the pipe still equipment, if depreciated in 10 years, amounts to about 0.82 cents per barrel.

Treating Costs. If the fractionation is poor and considerable entrainment occurs, the amount of acid required for off-color stocks may amount to as much as one pound per barrel. This cost has not been charged to the shell still battery in the summary which follows but it may amount to 0.2 cents per barrel.

Supervision, Water, and Overhead. These items should be much the same for either type of equipment. However, constant repairs take much of the time of supervisory managers and hence the supervision will be somewhat higher for the shell still battery.

SUMMARY OF COSTS

The above factors are listed in Table 3. Shell still I refers to a battery that has been partly modernized, whereas shell still II refers to batteries that are in poor condition with respect to repairs, equipment, and depreciation.

These costs are the ones that can be directly charged to the distillation operations and do not include plant overhead, storage, laboratory, sales, executive, taxes, treating or handling costs.

TABLE 3
Summary of Pipe Still and Shell Still Costs

	Cents per Barrel of Crude Oil		
	Pipe Still	Shell Still I	Shell Still II
Inadequate fractionation	0.0	0.7	1.7
Steam	0.6	1.47	1.9
Fuel	0.88	1.2	1.47
Still operation	0.72	1.08	1.3
Repairs maintenance	0.50	1.56	2.4
Power	0.03	0.04	0.03
Depreciation obsolesence	0.82	0.71	1.55
	3.55	6.76	10.35
Loss due to shell still operation:			
Cents per barrel	....	3.21	7.80
Dollars per year	....	$35,100	$85,300

Thus, the operation of shell still battery will involve a loss of about $35,000 per year. The figure may be as low as $20,000 a year for wisely operated shell still batteries but the cost may reach $85,000 per year on old, upkept equipment.

The above figures are for the distillation equipment alone and as a general rule those plants that operate shell still batteries also operate other equipment that is obsolete. For the complete operation of skimming including all direct plant expenses, the cost will probably exceed 12 cents a barrel for pipe still operation and from 18 to 24 cents for shell still operation.

Design of Continuous Treating Plants

IN the accompanying article which Mr. Trusty has contributed as one of the series on the general subject of Engineering Economics, the design construction and operation of continuous treating systems is discussed. The author presents data on types of mixers, such as orifice plates, baffle plates, jet mixers and mechanical mixers, and discusses pumps, settlers, blow cases and the size of pipe lines used with the system.

ENGINEERING ECONOMICS

A. W. TRUSTY
Chief Chemist, Louisiana Oil Refining Corporation

THE conventional method of treating distillates in the early period of refining was by the batch agitator (Figure I*). Contact was obtained by either circulating chemical and oil with a pump or by blowing with air, or both. The disadvantages were that air agitation entailed large evaporation losses, gave inefficient contact between the acid and oil, impaired the quality of the products in some cases, and, in order to get the desired results, an excessive amount of acid was used.

Batch operation was also undesirable due to loss of time while the agitator was being charged and pumped out. The modern continuous treating system has many advantages over the batch agitator:

(1) Much time is saved due to a continuous operation;

(2) Evaporation losses reduced considerably due to a closed system under pressure, and to the elimination of air blowing;

(3) Savings in chemicals due to more efficient contact of oil and chemical.

Chemical treatment of oil involves two fundamental principles: (1) efficient contact between the oil and the reagent, and (2) separation of the oil and reaction products. Efficiency of contact implies both efficient mixing and adequate time of reaction.

TYPES OF MIXERS

The types of mixers used in continuous treating plants may be classed as

1. Orifice columns;
2. Baffled columns;
3. Packed columns;
4. Mechanical mixers.

ORIFICE PLATE MIXERS

The orifice column has proved a satisfactory type of mixer that can be constructed on the plant. Methods of construction are shown in Figure II. The orifice area should be divided into several small holes rather than large ones. The energy of the high velocity jets is

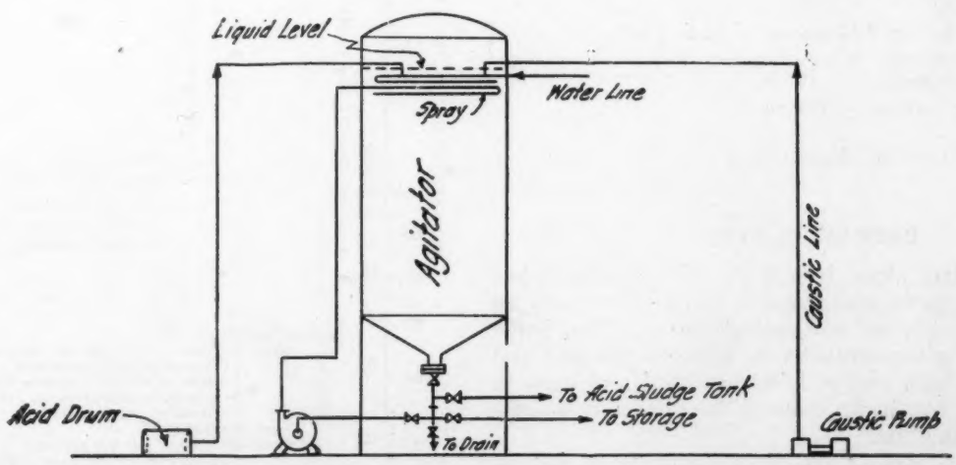


FIGURE I
Agitators with circulating system.

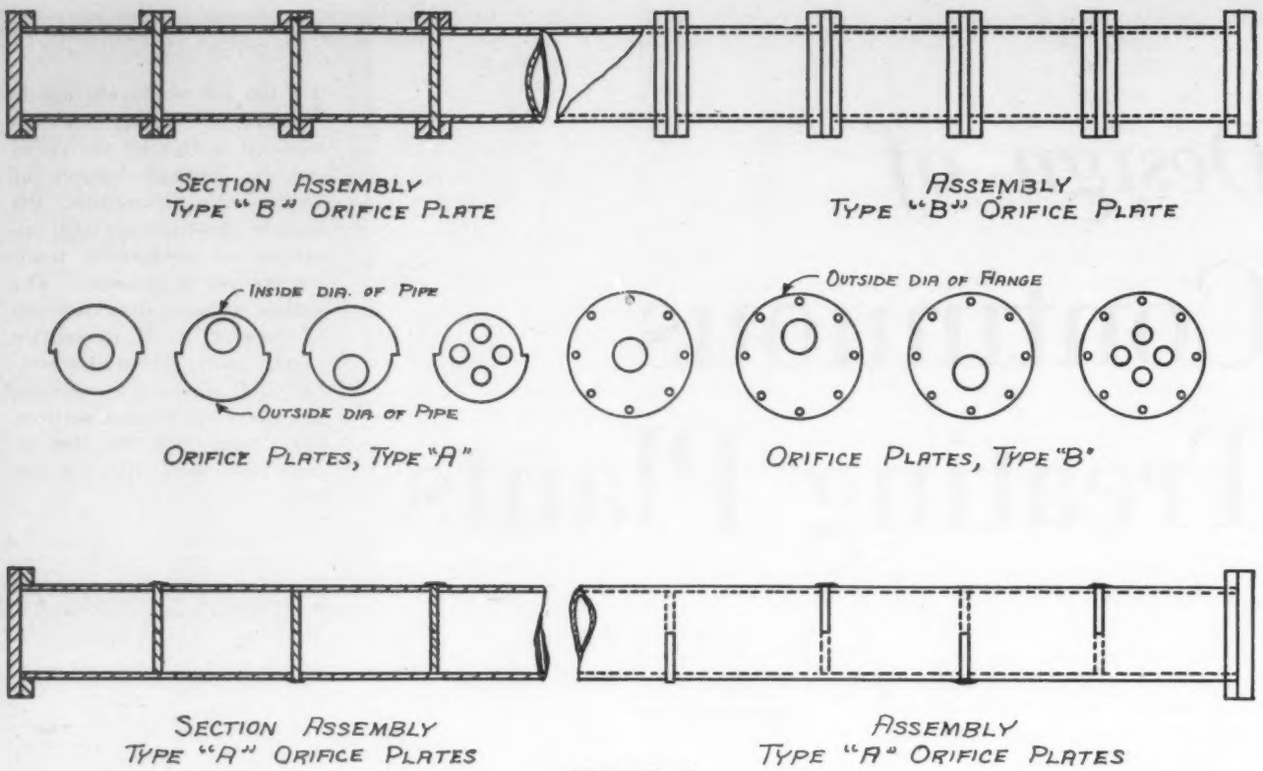


FIGURE II

Construction details for two types orifice mixers.

dissipated on the downstream side in eddies and turbulent motion which gives efficient contact.

The conventional orifice formula $Q = CA\sqrt{2GH}$ may be used to design orifice type mixers:

Q = quantity in cu. ft./sec.
C = coefficient (0.61 for small diameter orifices)
A = area of orifice, sq. ft.
G = gravity acceleration = 32.2 ft. per sec. per sec.
H = head of liquid in feed.

An adaptation of this formula is more convenient for common terms:

$Qb = 18.26\ D^2\sqrt{H}$

Qb = Flow in barrels per hour
D = Diameter of orifice in inches
H = Head of liquid in feet.

Example: To design an orifice type mixer to treat 250 barrels per hour of 58° A.P.I. gravity distillate with a 30-pound allowable pressure drop. Using 15 plates, the pressure drop will be 2 pounds per plate.

$250 = 18.26\ D^2\sqrt{6.16}$ or

$D^2 = 5.52$

D = 2.35" or 4.34 sq. in of orifice area per plate.

Rather than have one or two large orifices, better mixing can be obtained with, say, ¼" holes.

Area of one ¼" orifice = 0.049 sq. in.

$$\frac{434}{.049} = 88.5 \text{ holes } \tfrac{1}{4}\text{" diameter.}$$

BAFFLE PLATES

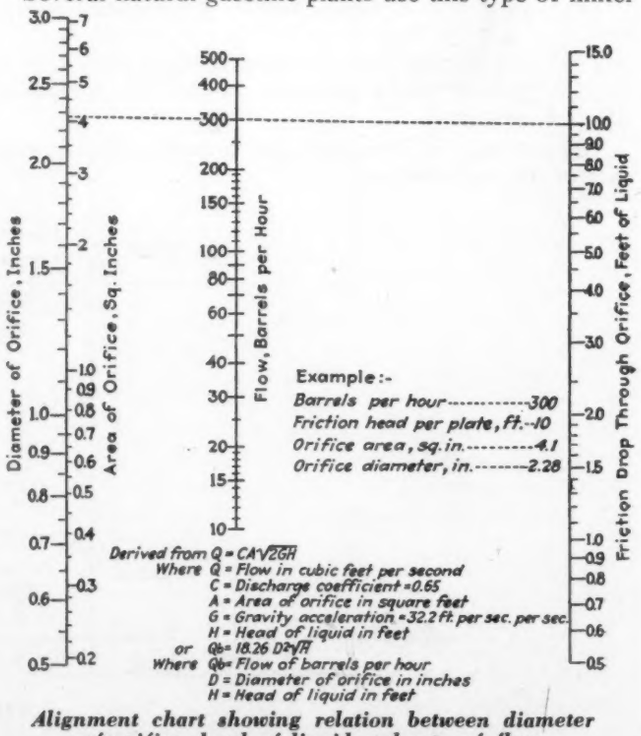
The baffle plate alone is not an efficient mixer, but can be used to good advantage in conjunction with an orifice column, jet, or mechanical mixer. The baffle serves to prolong the contact time between the acid and the oil. Where acid sludge is recirculated and used in large volumes, the baffle column can be used to advantage for contacting.

Twenty baffles spaced one foot apart in a six-inch column and treating at the rate of 200 barrels per hour will give a pressure drop of about 10 pounds.

JET MIXERS

A Venturi tube or a number of smaller tubes in parallel are sometimes used as mixers. A high velocity jet of liquid flowing through the throat of the nozzle creates a suction which draws in the chemicals. This type of mixer is most efficient when approximately equal volumes of liquids are to be mixed. Usually the quantity of chemical is small compared to the quantity of oil; hence it is difficult to positively control the relative amounts of liquids.

Several natural gasoline plants use this type of mixer


Diameter of Orifice, Inches
Area of Orifice, Sq. Inches
Flow, Barrels per Hour
Friction Drop Through Orifice, Feet of Liquid
Example:-
Barrels per hour----------300
Friction head per plate, ft.--10
Orifice area, sq. in.---------4.1
Orifice diameter, in.--------2.28
Derived from Q = CA√2GH
Where Q = Flow in cubic feet per second
C = Discharge coefficient = 0.65
A = Area of orifice in square feet
G = Gravity acceleration = 32.2 ft. per sec. per sec.
H = Head of liquid in feet
or Qb = 18.26 D²√H
Where Qb = Flow of barrels per hour
D = Diameter of orifice in inches
H = Head of liquid in feet

Alignment chart showing relation between diameter of orifice, head of liquid and rate of flow.

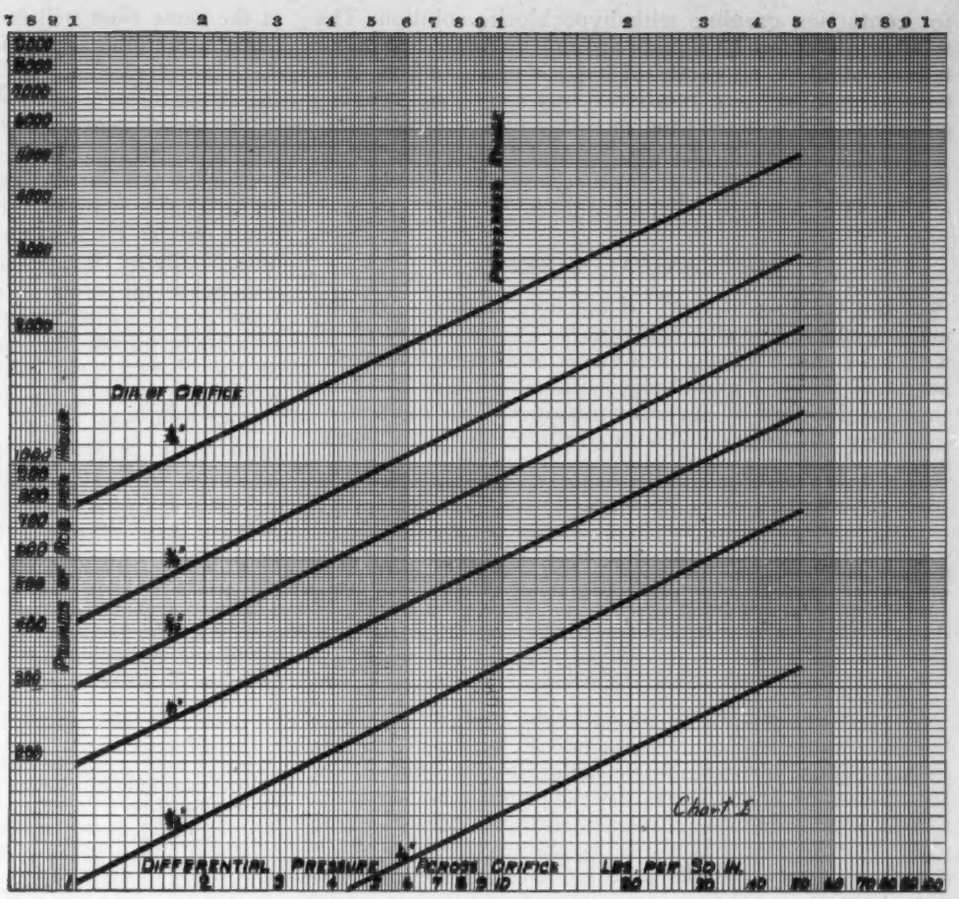


CHART 1

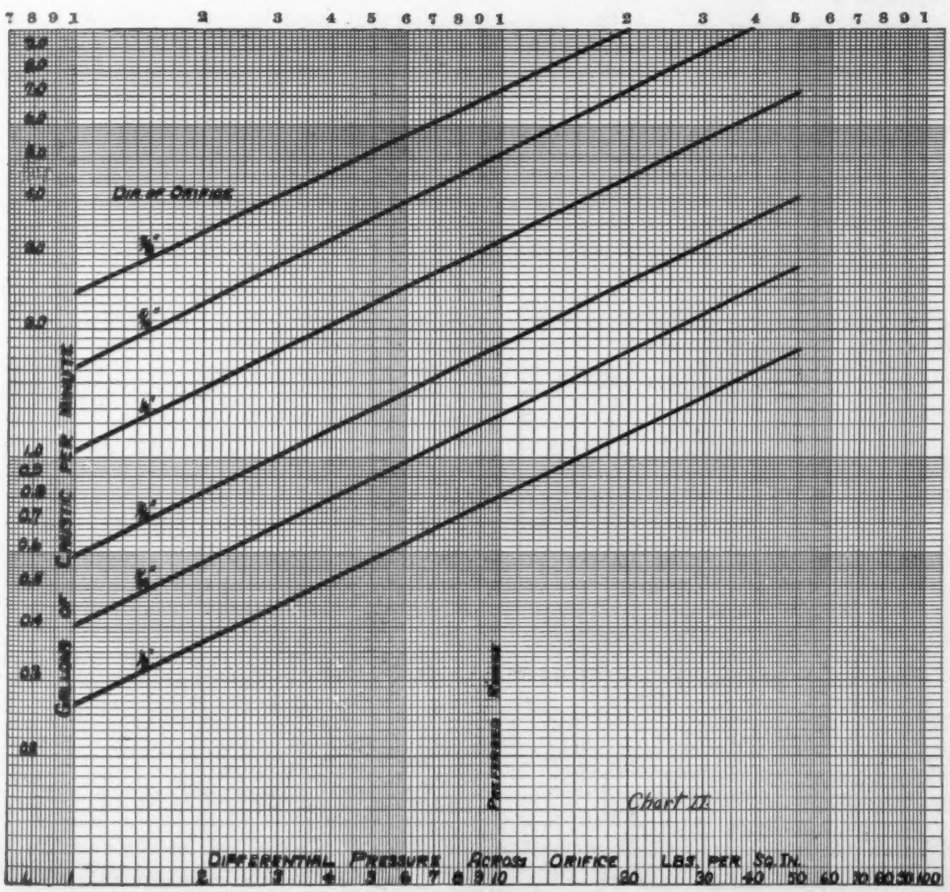


CHART II

for contacting gasoline with hypochlorite solution. The relative quantities of liquids are not important to maintain constant, as an excess of hypo is used, and, as most natural gasolines are relatively less sour than refinery gasolines, the contact time is sufficient for sweetening.

The packed column—a pipe packed with crushed rock, Raschig rings, gravel, etc., gives efficient mixing, but has the disadvantage of the tendency to clog.

MECHANICAL MIXERS

There are several types of mechanical mixers on the market which give efficient contact. The treatment of oil with acid with a mechanical mixer and immediate separation of the acid and sludge with centrifuges has been successfully used by several companies. This method of treatment has been dealt with considerably in the literature.

PUMPS

Centrifugal and rotary pumps have some value as mixers, but as a rule do not furnish sufficient contact to be used alone. Pumps assist the mixing when used in conjunction with some other type of mixer.

SETTLERS

The settlers should be of sufficient capacity to allow the bulk of the chemical to settle out. Settlers with low capacity cause an excessive consumption of caustic, or, if a water wash is used, the corrosion will be more severe on the water-wash settlers and lines. On the other hand, if the settlers are so large that excessive settling time is given, the quality of the treated distillate is impaired. Prolonged settling causes esterification, which gives excessive acid formation, a poor color, and a poor color stability when the treated distillate is rerun.

An average settling time of one hour is sufficient for most types of distillates. For a 300-barrel-per-hour plant, an 8 x 30-foot settler after the acid mixer and an 8 x 30-foot settler for auxiliary settling should be ample.

Normally, the inlet to a tank should be made at the point where the separating level is kept. If the inlet is made tangential to the tank, a whirling motion is set up which will assist the action of gravity in settling, and at the same time will not tend to cause as much turbulence in the tank as an inlet straight into the side.

BLOW-CASES

The most convenient method of handling sulphuric acid is by means of air and blowcases. The quantity of acid used can be measured by the differential pressure on an orifice plate in the acid line (see Chart I) or by down gauges from gauge glasses on the blowcases taken periodically by the operator. When using an orifice plate, a monel metal screen should be placed in the acid line in front of the orifice to prevent clogging. The proportion of acid to oil can be changed by changing the pressure on the blowcase or by substituting an orifice of different size.

The blowcase should be designed to withstand the refinery air pressure after an allowance has been made for corrosion. An allowance of one eighth-inch should be made for corrosion. Check valves in the air line will prevent acid leaking back in case the air supply is shut off.

SIZE OF LINES

The velocity of the oil through the various lines should be approximately five feet per second. A convenient formula used to calculate this velocity is

$$\frac{.2856 \times \text{barrels per hour}}{(\text{diam. of pipe in inches})^2}$$

For example, the size piping to use for 250 barrels per hour would be

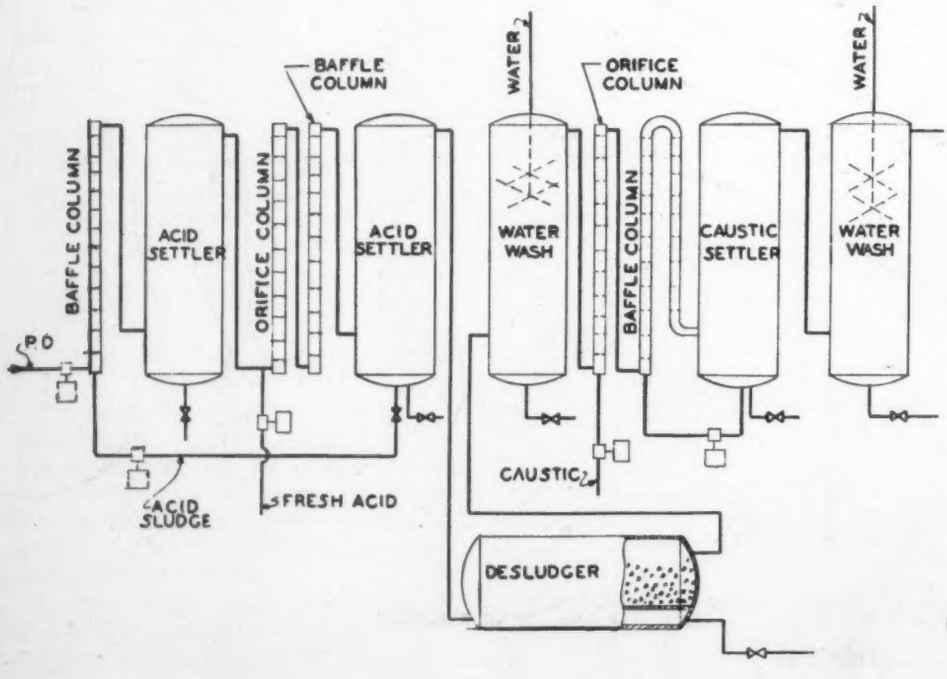
$$\frac{.2856 \times 250}{D^2} = 5$$

$D = 3.8''$. Use 4" pipe.

DISPOSAL OF ACID SLUDGE

The method of disposal used by many refiners is to burn the acid sludge. The products of combustion of the acid in the sludge, sulfur dioxide and sulfur trioxide, are non-corrosive, if not allowed to combine with water. Where disposal can be made into nearby streams, the sludge should be neutralized with spent caustic.

References:
*Morrell & Bergman, Chem. & Met. 1928, 35 Motor Fuel Prep. & Application—Nash, Howes.

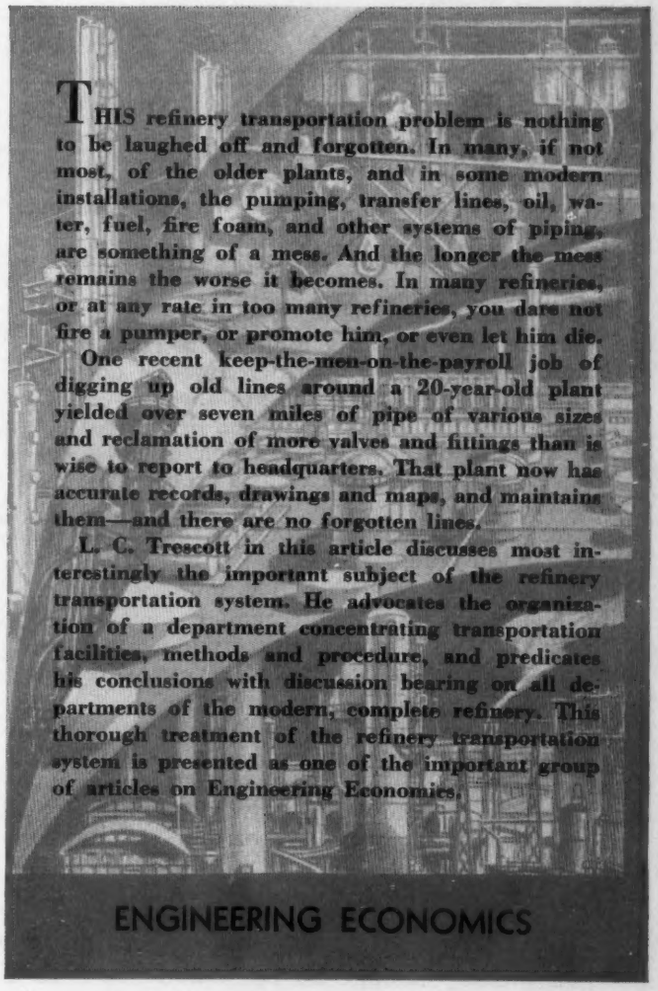


Continuous treater for using acid sludge as first treat.

THE REFINERY

Transportation System

By L. C. TRESCOTT



THIS refinery transportation problem is nothing to be laughed off and forgotten. In many, if not most, of the older plants, and in some modern installations, the pumping, transfer lines, oil, water, fuel, fire foam, and other systems of piping, are something of a mess. And the longer the mess remains the worse it becomes. In many refineries, or at any rate in too many refineries, you dare not fire a pumper, or promote him, or even let him die.

One recent keep-the-men-on-the-payroll job of digging up old lines around a 20-year-old plant yielded over seven miles of pipe of various sizes and reclamation of more valves and fittings than is wise to report to headquarters. That plant now has accurate records, drawings and maps, and maintains them—and there are no forgotten lines.

L. C. Trescott in this article discusses most interestingly the important subject of the refinery transportation system. He advocates the organization of a department concentrating transportation facilities, methods and procedure, and predicates his conclusions with discussion bearing on all departments of the modern, complete refinery. This thorough treatment of the refinery transportation system is presented as one of the important group of articles on Engineering Economics.

NOT so long ago a department head in one of the largest refineries in the East walked into his main pumphouse.

"Tommy," he remarked significantly to the pumper, "I wish we could lose that mess in 107 tank. It'll never make cold test. It's too sloppy to press and they won't let us throw it into fuel. As long as we have it they'll think there's a chance of working it up and we need the room."

The next day 107 tank was empty. About a month later an old tank in the oldest tank farm about two miles down the river was found full of sloppy red oil. How it could possibly have reached the tank no one was able to explain. On the face of it part of the trip must have been made straight through the air. After a week or so of wonderment they threw the stuff into fuel. The secret of its passage was in all probability buried with two old pumpers who had worked in the refinery when it started. They had been carried on the payroll for some years after they should have been pensioned just because there were combinations of old lines which they alone knew.

The pumper in that department also makes his brag that you cannot drive a stake within a hundred feet of his pump house without hitting a line. He does not brag of the fact that many of them are forgotten lines; lines that tangle in and out among themselves; lines that begin nowhere and for all practical purposes go nowhere. A tie-in or cross-over costs many times what it should because of the necessity of weaving a way through. Of this tangle too often the only surface indication is a thick sprinkling of valve wheels and short projecting pipes which house the valves of the lower lines. To the new man, trying to learn the lines looks like a lifetime job, and frequently that is just what it is.

A line may start as a feeder for one or two tanks and in the course of 10 years be extended and tied in until through it almost any tank in that section of the yard may be reached. Quite handy, you think? Remember that for a single pumping on that line every valve must be tried. The pumper may walk a mile in order to pump a tank within 50 feet of the pump house.

Such conditions do not hold for every refinery. Some are still young. But with the best of them, unless some care is taken to prevent it, inside of 20 years the pipe lines can and do reach the stage where they can be compared more nearly to the aimless, tangled trails of an age-old jungle rather than to a planned transportation system. In any plant modernization scheme it would seem a worthwhile project to tear apart at least some of the mess into which the lines have gotten themselves and rebuild with less complication. In any event there should be some record of the lines which possibly may be put in use and this record should not be kept in the heads of a few of the older pumpers.

When a plant is built the tanks and equipment are located and after this the connecting pipe lines are put in. Seldom indeed is the matter of piping and pumping considered in any other light than just the inter-connection of tanks or equipment. Seldom does the engineering department look at it by itself, as a matter of transportation with problems of dispatching, extension, maintenance, and sundry difficulties all bearing on one point: that of keeping the crude

and its products moving in a smooth uninterrupted stream.

Usually the questions considered by the engineering department are confined to the selection of proper pipe sizes and to the original location of the lines. Subsequent extensions and connections are often laid out by the operating foreman to suit the needs of the moment. Viewed as a major problem of transportation the proposition is more spread out. Here are some of the items which must be considered:

1. Location of tanks.
2. Location and layout of pipe lines.
3. Quantity and quality of oils to be transported.
4. Sizes of pipe lines and the purpose for which each line is to be used.
5. Selection of proper pumps to meet each condition.
6. Proper dispatching of pumpings in order to avoid hold-ups and congestion.
7. Special precautions necessary for each case.
8. Prevention of contamination.
9. Careful consideration of all changes and extensions of pipe lines.
10. Safety considerations; fire hazards, firebanks, etc.; foam lines, and safety steam lines.
11. Transportation of reagents, acid, soda, etc.
12. Transportation of acid sludge.
13. Records. Source data for yields.

Some of these items are highly important, others are not, but they can all be drawn together under the head of refinery transportation. Possibly the very fact that transportation is so closely tied in with all plant processes has been the thing which in most cases has prevented its being treated as a problem by itself. In any event it is certain to be of value to separate this transportation network from the rest of the plant operations and study it.

The flow sheets, Figures I to V, give the transportation layout for a complete refinery, one running lubricating, coking, and asphaltic crudes. The smaller refineries can easily pick out the lines in which they are interested.

It will be seen that the heart of this transportation system lies in the pump houses of the distillation department. Here the crude is received, the primary split-up made and the various raw stocks dispatched to the proper department for further refining and finishing.

The operating tank farm usually comes under the jurisdiction of the distillation department. This farm has two functions in the transportation system. It provides primary storage for crude and it acts as a storage balance for the plant as a whole. While, in the usual run of things the various stocks are transported directly from the tanks of one department to another, there are times when holdups occur. The regular stream becomes clogged and in order that the stills and the plant may be kept moving the excess stock must be sent to the farm.

This use of excess storage is not always caused by emergencies. In the case of fuel oils, certain lubricating stocks, etc., the requirements of the trade are seasonal or intermittent. The shipments may be in large cargoes and not in a steady stream. Reducing stills are run for a varied line of products. Usually they are kept on one line until a considerable stock is built up. Provision must be made to store this until it is needed. It is usually better policy to store the raw stock rather than the finished product so that such oils generally are held at the farm.

Pumpings to and from the operating tank farm are usually in large quantities. A careful study of the actual needs should be made and the pumps and lines laid out accordingly.

The dispatching problems of the tank farm are relatively simple as the daily pumping routine is fairly constant. And as one purpose lines are almost the rule there is little complication due to handling several stocks in the same line.

The first line tanks in the distillation department are naturally the still running tanks. Here are the feed tanks for the crude stills and the raw stock tanks for the reducing stills. In up-to-date practice for lubricating oil manufacture the coastal and naphthene base lubricating stocks are cut to specification on the crude still, so that for these grades no further reduction is necessary. The reducing stills generally will be found operating on pressed distillates either from straight-run crudes or from the coking stills.

At least two feed tanks should be provided for each stock or crude. With several continuous stills operating on the same stock there should be a feed tank for each still plus one or more extras for the system. When batch stills are used several may be charged from the same tank. When a feed tank is empty it is immediately filled from the tank farm, the wax plant or other sources, the still run being continued from one of the extra tanks. Each tank should be large enough to hold at least a day's run for the still or stills which it feeds, and direct one purpose line should be laid between the feed tanks and the source of supply. The pumps should be of such a size as to re-fill the tanks as soon as is reasonably possible.

The running tanks from the stills should not be too large. Two thousand barrels is about right to handle the product from a still of 6,000 to 10,000 barrels per day capacity. There should be two of these for each cut of oil from each continuous still, or from each crude battery of this capacity. The continuous tower still will also have a tank for reflux circulation. This is pumped occasionally to keep it at a proper level, the product going to cracking stock.

Back of the still-running tanks are the larger storage tanks into which the running tanks are pumped after they have been passed by the laboratory. Naphthas and W. W. distillates generally are stored in the refined oil department as are the finished products from such stocks, the gasolines, the kerosenes, etc. In this way the light storage is segregated to reduce fire hazards. The cracking stock storage generally is spread between the distillation department and the cracking stills.

Where paraffin-base crudes are run the most important and most valuable are the bottom stocks or residuums. Fresh from the stills these stocks carry the petrolatums and the tanks in which they are stored must be kept hot. They generally are run direct from the stills to the larger storage tanks without going through the smaller intermediate tanks, and in laying out the system for them a careful study should be made of the situation. The running rates from the stills must be considered as well as the rates at which the stock will be drawn by the various treating departments. Plenty of allowance must be made to take care not only of possible holdups on the treating end but also to allow for such periods as the stills may be shut down for repairs. In addition to all the tank space that may be considered necessary to handle these grades of

oil it is well to install at least one extra, of considerable capacity. It is surprising how often it will be used.

Asphalts usually are run and held in a department of their own. They too go direct from the stills to storage in heated tanks. Short lines, often jacketed, are the rule and every precaution must be taken to keep the product dry. With many stocks a leaky coil may only mean so much wet oil. With a heavy tar or asphalt it means foam without end. A very little water in such a tank will practically empty it through the manhole at the top and the chances of recovering the solid tar from the ground or of opening sewers that are plugged with it are slim indeed. In many plants it is against the rules to blow heavy tar or asphalt lines with air on account of the moisture that may be carried by the latter. Welded heating coils should be the rule in such tanks.

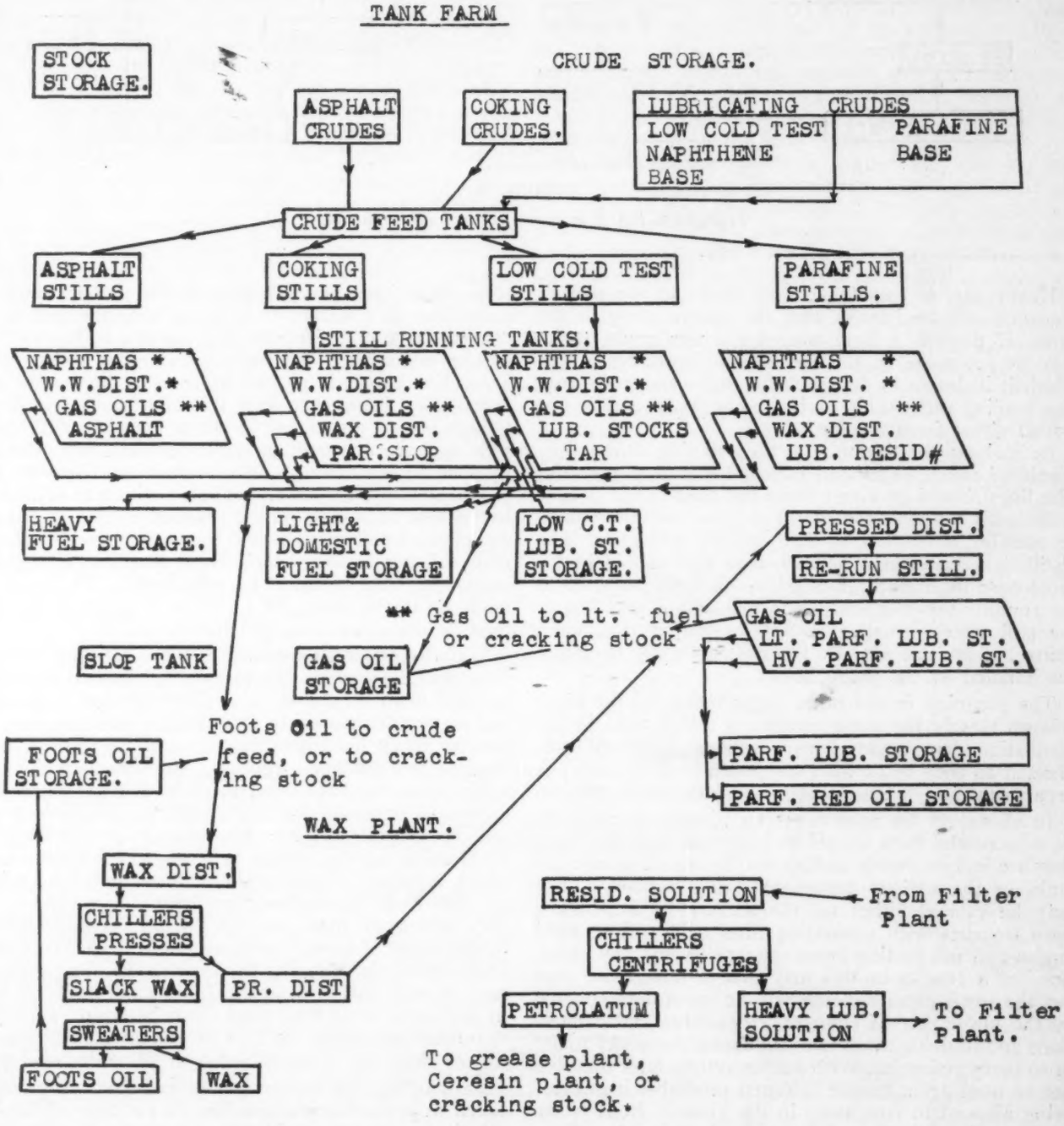


FIGURE I

Transportation flow sheet, distillation department

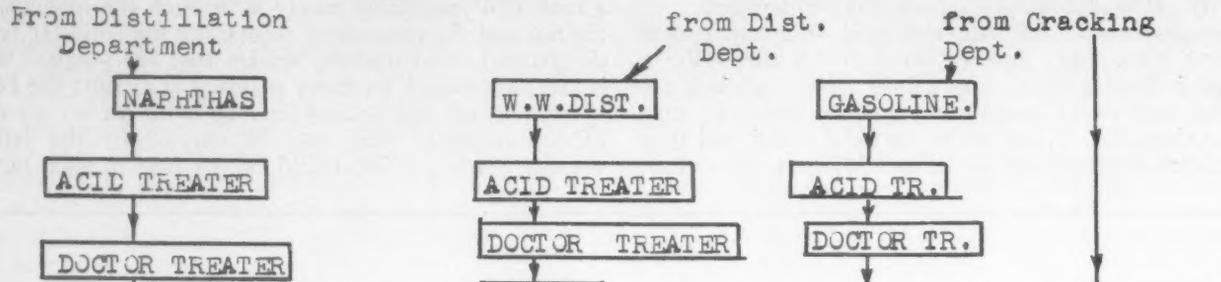


FIGURE II

Transportation flow sheet, refined oil department

Heavy fuel is usually pumped direct to the power-house or still feed tanks with the excess going to the farm. If there is a local market for this grade, tanks may be set aside in the distillation department from which it is drawn to the trucks or tank cars at the regular loading racks. Ship or barge cargoes generally are loaded direct from the farm.

In making the layout for the running tanks they should of course be located as near the stills as possible. The lines should go direct from the stills to the proper tank and it is almost the law of the land that no outside or transfer lines shall be tied into the still lines. The distillation department pump houses also should be located near the running tanks. The oil from the stills to the running tanks is handled by the pumps attached to the still which usually are not located in this pump house but are operated by the stillmen. Only transfers are handled by the pump house.

The pumping in the other departments of the plant follows closely the same conditions which hold in the distillation department. Their problems are less complicated as they work with the products from that one department and their operation is pretty much routine.

In all except the most northern climates there is little reason why lines should be buried, especially where they are laid as closely as they will be found around the tanks of the various departments of a refinery. They may be carried either on the surface or in shallow open trenches with connecting lines going off at right angles and not cutting cross country in just any direction. In a few cases this may take a little more pipe but the small extra expense will be more than covered by the lower cost of changes, cross-overs and connections and in doing away with the messy networks found in so many refineries. With such a system lines that will not be used again can be salvaged profitably instead of being allowed to rust away in the ground. Repairs are made easily and plugged lines can be opened without the expense of digging them up. There is also a great advantage in having everything in sight. Lines can be followed and a pumper has a fair chance of learning all of his lines before he is ready for a pension.

In the beginning many layouts are made in this way with the best of intentions of holding to the original plan, but as the years pass new lines are thrown in under the pressure of one emergency or another, and trick connections are put in until the pumper must be something of a wizard to keep up with the puzzle.

With large organizations it is possible to have a part of the engineering department take care of the pipe lines. Every line is plotted as soon as it goes in and every effort is made to keep this map up to date. Even though such a proceeding is not followed it still is possible to have one man in the department who has the lines under his care. He is consulted on all extensions, cross-overs and tie-ins, and is in a position to hold them to the scheme laid out in the beginning. Large maps are complicated and cumbersome for general use but sketch maps of the various single lines, showing all connections, etc., can be made of great value. They can be made up in the same size as the tank gauge sheets and kept in the same sort of binders.

With long direct lines such as those between the tank farm and the various departments it is not so necessary to hold them on the surface although such a practice has many advantages. In some of the more modern refineries trunk lines have been carried in tunnels or in shallow concrete-lined trenches. Any scheme which gets away from the haphazard methods heretofore used is certain to be worth while. In plant modernizations and making extensions some consideration should be given to clearing up the tangle of lines which exists in so many refineries. Get the unused lines out, where it is possible to do so, and salvage them. Their value probably will more than pay for the job. And in putting down a line it is well to figure that some day it may come out again. Keeping the lines in the clear makes it easy to take them out again. Unless the lines are kept in the open or at least kept from becoming a part of the usual bad tangle the cost of salvaging them may be higher than any possible value of the recovered pipe.

In handling the various pumpings some care must be taken to prevent contamination. In the case of finished oils it is necessary to have a separate line for each grade. In the distillation department a slight contamination is not a matter of any great importance except on such oils as must meet a flash specification. Practically everything going from this department must be further treated or refined. Naphthas of various grades may be handled over the same lines and by the same pumps from the running tanks to the refined oil department. Water white or kerosene distillates may be han-

dled together. Various grades of light lubricating stocks may go over the same lines, due attention being paid to flash. The heavy residual oils are sent over their own set of lines. Care must be taken that no lighter oils get into these lines as flash is important here. Tars must of course be kept in their own system but all of the lighter grades of fuel may be handled over the same trunk lines. In the case of fuels or other stocks which must be protected against wet or dirty lines, or fuels which are being pumped for shipment, provision should be made to circulate through the entire length of the line until the stream is satisfactory. Usually such circulation is from fuel storage through the lines and the pump and into cracking stock. Care must be taken that wax distillate and pressed distillate are kept apart. Often the so-called regular pressed (cold test about 20° F.) and the zero pressed are kept in separate lines. Under some conditions, however, when the pumpings are large the same lines may be used for both.

When cold weather comes on, steam must be put on many of the tanks. For one reason or another steam is kept on many tanks all year round. The heavy oils and tars must be kept hot in order that they may be pumped. As stated before, the coils in such tanks should be of the welded type as leaks may cause foaming from wet oil.

Heat is also required on many oils that must be freed from water, as fuels, etc. For this purpose it would seem that some sort of dehydrating equipment such as a centrifugal spray would save considerable time and steam. With such equipment the oil could be pumped through an exchanger for heating then on through the spray where the water would be evaporated. The oil going into storage would be clear and dry and the heating would be done under the most efficient conditions. In most cases the oil is pumped into large tanks and heated by means of bottom coils to a temperature running from 120° F. to 180° F. The water does settle out but all this takes time and the coils are not particularly efficient as heaters.

Another winter job required of pumpers is to see that all valves and valve boxes are properly protected against freezing. Spent percolation filter clay makes an excellent packing for valve boxes.

It is not always fully appreciated by all concerned that the source of all yield figures are the agencies handling the transportation of the oil in the refinery. The process yields, the still percentages, etc. are all taken from figures derived during the gathering and transfer of stocks. The commonest record is the pumper's log sheet. The form of this is more or less standard the world over and is shown in Figure VI. On this log are entered not only the transfers but a record of all activities of the pump house. It not only shows the stocks transferred but also shows what lines are in use. Often there are valves on through lines which are located in the area covered by the pump house or certain valves must be closed or "bucked" while other departments are pumping. These operations all are shown on the log. Some concerns take all of their figures direct from these sheets. Others use in addition to the log a transfer sheet. This sheet is practically a duplicate of the log in form but in addition to transfer figures, gauges, etc. of the pump house itself it is the practice to call the other end of the line after a pumping and enter their figures for the oil transferred. Any

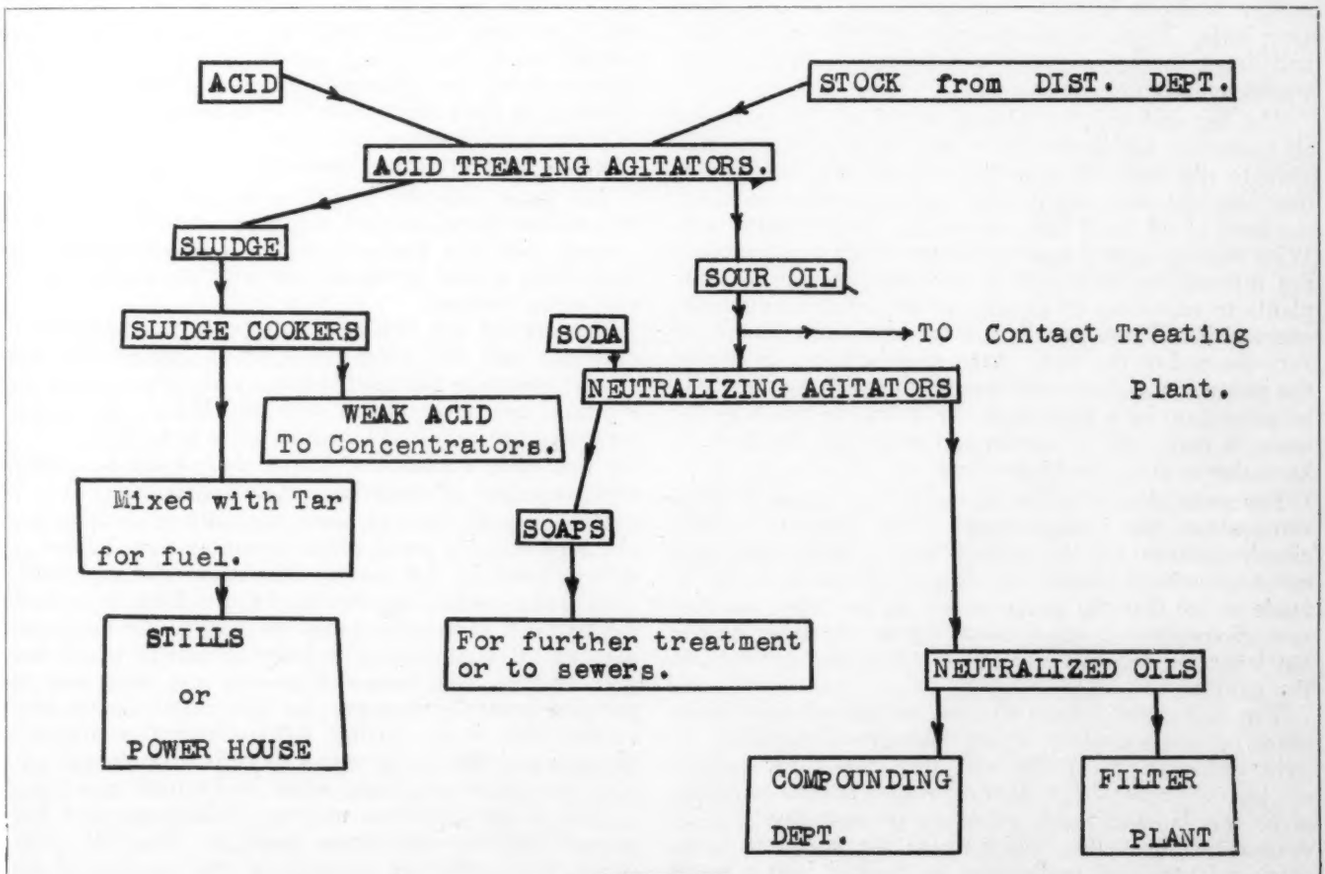


FIGURE III

Transportation flow sheet, acid treating plant

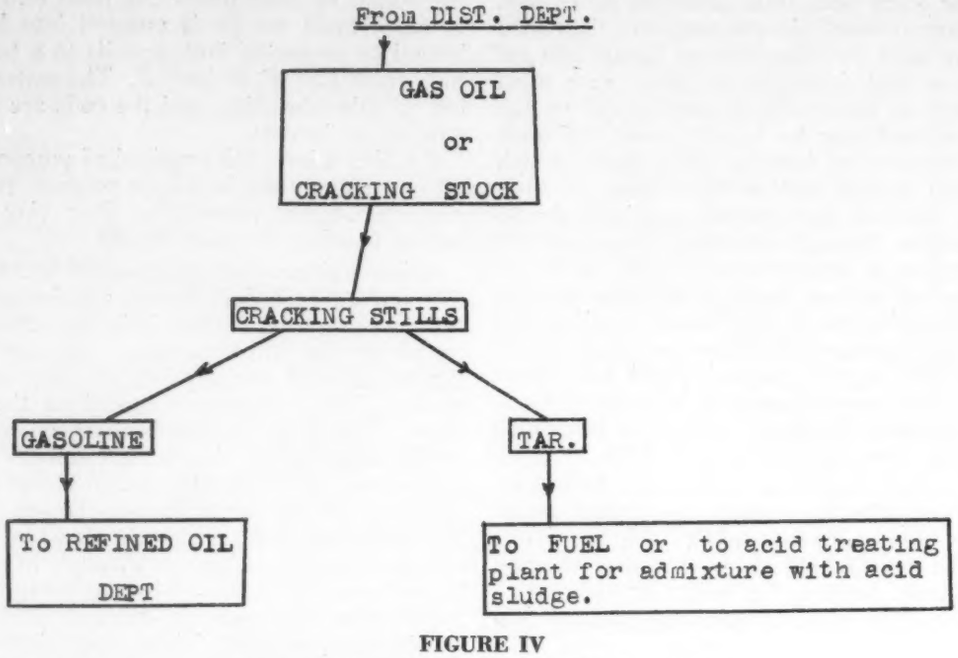


FIGURE IV

Transportation flow sheet, cracking department

difference in the two sets of figures is noted and is either checked and corrected or explained. Such a practice is good. It enables an immediate check on any differences by the men who did the work while the whole matter is fresh in mind and the tanks have changed very little. These transfer sheets are sent to the office and from them the yields are taken. Several carbon copies usually are made.

The log and transfer sheets record all transfers but all tanks are not in use every day. It is of course possible to dig back through the log and get any figures that may not show up at once, but most concerns check the level of oil in all tanks every day, at or about 7 a.m. With these gauges temperatures are taken and the tanks are thieved for water. It is also the practice in many plants to run a set of gauges on all still running tanks every shift. These usually are made about an hour before the end of the shift. They may be taken by one of the pump house force and sent to the stills or they may be picked up by a man from the stills. In this way the room in each tank is known and plans may be made to keep the various stocks moving.

For general refining work the outage gauge is used rather than the innage gauge. This method is sufficiently accurate for the purpose and is much faster and more convenient to use. An occasional check should be made to see that the gauge points on the tanks are correct. Sometimes a cover may slip or the edge of the hatch may become so worn as to affect the accuracy of the gauging.

The only point where the use of special type tanks must be considered is in the storage of naphthas or light products or in the still feed tanks for certain grades of crude. Here floating roofs, special vent devices can be used with profit to prevent loss due to evaporation or to the breathing of the tanks. In some cases a battery of tanks may be hooked into a small recovery system through which the tanks breath.

Practically all tanks are provided with safety steam. This is carried to the tanks from the regular steam lines. The control valves are placed in manifolds at some distance from the tanks and in a position where the steam may be put on without danger to the operator. Care should be taken that all concerned are thoroughly familiar with the location of the valves and the application of the steam to the tanks. These lines usually are blown out and tested twice a year. In maintaining them the principal difficulty comes from the splitting of lines and valves due to freezing. It is hard to prevent small drips, and often old valves will leak. In freezing weather some attention should be paid to this.

The same difficulty is met with foam lines. In addition to this the aluminum sulphate solution often picks a weak spot in a line and chews its way through. The foam lines should be blown out with air about once in every two months.

The special fire risk is of course in the handling of naphthas and the lighter crudes. Stocks of this sort should never be pumped during a thunder storm and the lines are never blown with air. Many other safety measures are required by underwriters, by law, and by the operating companies. All of these must be considered as a part of the transportation problem.

Except in the case of heavy asphalts or of some special high melting point waxes, steam jacketed lines are seldom used in the normal operation of the refinery. The tanks containing the stocks are held at a sufficiently high temperature that the oil may be kept moving once it is started. For heavy products which may give trouble high pressure pumps are used and the pumper generally watches his job closely until he is certain that it is moving right. Once the stream is through and the line is warm there is little danger of a plug, but there are times when really high pressure is needed to get the lines moving. With long wax lines connections are sometimes made at intervals where steam hoses may be hooked on and a plugged line opened by this means. In cold weather every care must be taken that all lines which may give trouble are blown with air before they are shut down. Lines are often

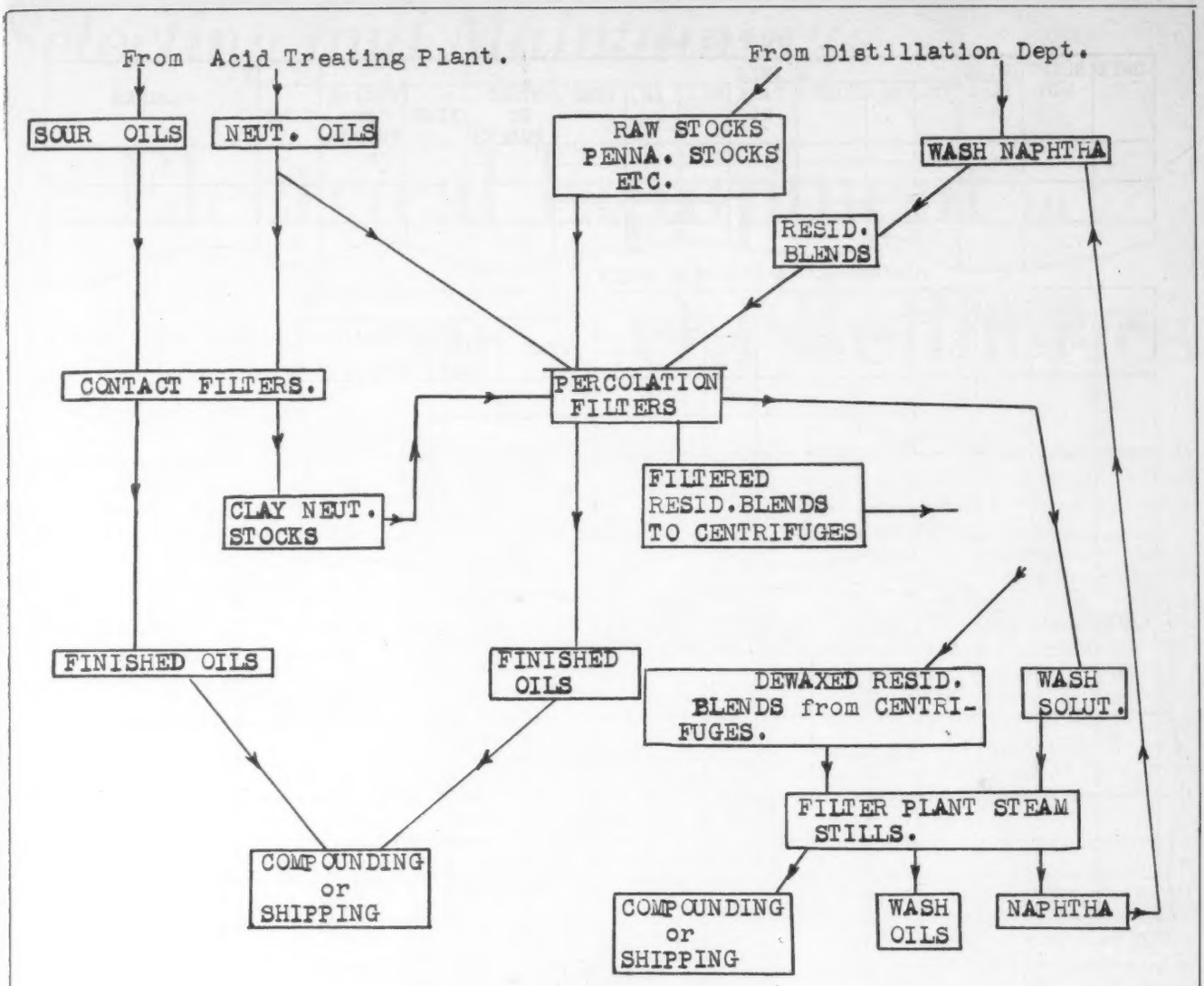


FIGURE V

Transportation flow sheet, filter plant

plugged due to a leaky valve or to the line being left open by mistake. In cold weather only constant vigilance will keep the lines open in which tars or waxy oils are handled.

In pumping acid of the strength generally used for treating, iron or steel pipe may be used. This also applies to soda or other reagents. A large portion of the acid will come from the acid recovery plant. To keep this up to strength and to replace losses, fresh acid will be brought in, but in neither case does this involve any special difficulties except the usual safety precautions. Pumps of the type used to handle acids in chemical plants may be used in the acid plant, but ordinary type pumps are used in the treating plant. In handling weak acids or other corrosive reagents regular chemical plant practice may be followed.

Soda may come into the plant as a solid in drums, in which case it must be dissolved before use, but the usual method is to bring it to the plant in tank cars. The usual strength of this solution is 44° Be., which is diluted with water to any strength desired.

Acid sludge to the sludge kettles is pumped through very short lines by specially designed pumps. The handling of acids, sludges, etc., has been fully treated in a previous article (REFINER, Vol. 12, No. 12, December, 1933).

The method by which acid sludge is pumped to the power house and stills for use as fuel will depend to a great extent on the sludge itself. Some may be mixed with lighter fuel, others will not dissolve in this at all. In most cases sludges can be transported well enough if they are mixed with a fairly heavy tar. This may be pressure still tar, still bottoms, or any tar of like nature. It has been found advisable to use brass lines for sludge.

In the refinery there is a very considerable amount of dispatching to be done even though it is not always recognized as such. Running tanks must be kept cleared; pumpings must be arranged so that there is no conflict in the use of lines; inventories of raw stocks must be kept up and finished stocks must be on hand for shipment. In each pump house the pumper must keep in mind all the lines in use in the area for which he and his helpers are responsible. He must watch his tanks, see that they are held at a proper temperature; that the running tanks are changed over before they run over and that they are pumped promptly. Feed tanks must be kept filled. The pumper is constantly called upon to find

PUMPERS LOG

DATE	SHIFT NO.	FROM TANK NO.	FT.	IN.	TEMP.	INTO TANK NO.	FT.	IN.	TEMP.	START BY PUMPER	TIME	FINISH BY PUMPER	TIME	REMARKS

PUMPERS LOG & TRANSFER SHEET.

PUMPING FROM

DATE	TIME START	TANK NO	OIL	WATER FT.	WATER INS.	GAUGE FT.	GAUGE INS.	INNAGE FT.	INNAGE INS.	TEMP.	GALLONS TANK TEMP	GALLONS @ 60°F.

PUMPING INTO

DATE	TIME FIN.	TANK NO.	OIL	WATER FT.	WATER INS.	GAUGE FT.	GAUGE INS.	INNAGE FT.	INNAGE INS.	TEMP	GALLONS TANK TEMP.	GALLONS @ 60°F.	DIFF. ON REC. TANK OVER	DIFF. ON REC. TANK SHORT	REMARKS

FIGURE VI

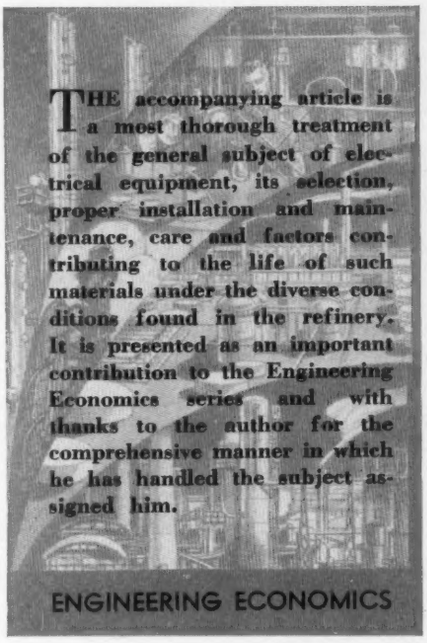
extra lines to get from one place to another and this is often complicated by the fact that the lines are buried and are not easily followed out. As a matter of fact a far greater responsibility rests on the pumper in a large pump house than is generally recognized. And it must be noted that all of his problems, while intimately connected with the department in which he operates, are essentially the problems of the more general matter of transportation.

This is not intended as a treatise on pumping. Most of the methods touched on are well established and well known but they are usually thought of and worked out only as a part of the operations of a particular department. It would seem a distinct step toward modernization if the transportation facilities of a refinery were concentrated in a department of their own, with someone at its head whose job is to study the whole situation and keep it properly coordinated. This man's work should not only cover the pumping of the various stocks but should also include the planning and layout of piping changes, extensions, etc., proper maintenance, pumping schedules, and the thousand and one other items which must be brought together to properly organize the transportation system.

Selecting and Maintaining

Electrical Equipment *in* Oil Refineries

V. R. CHADBOURNE


THE accompanying article is a most thorough treatment of the general subject of electrical equipment, its selection, proper installation and maintenance, care and factors contributing to the life of such materials under the diverse conditions found in the refinery. It is presented as an important contribution to the Engineering Economics series and with thanks to the author for the comprehensive manner in which he has handled the subject assigned him.
ENGINEERING ECONOMICS

IN planning a new installation it is always well to keep in mind that maintenance is one of the major items of production costs. Consequently all equipment selected should be studied thoroughly to be assured that this cost is kept within reasonable limits.

First cost can not be held to be the criterion of a good job but rather the cost of operation with its attendant upkeep. Simplicity of design in the process and equipment used is worth striving for. Although refinery processes are becoming more and more complicated yet the apparatus used need not necessarily reach a state that it is bewildering to the average repair man. It is still possible to make selections of electrical equipment that will meet the demands of modern processes and hazards and at the same time be easily maintained by average mechanics.

All industries have their own peculiar hazards and annoyances and the oil industry is no exception. Explosive vapors and volatile oils require a more or less special technique in the installation and upkeep of electrical equipment used in oil refineries, while the corrosive effects of the sulphuric and sulphurous vapors make desirable the use of oil-immersed or wholly enclosed controls and the use of protective plating on exposed contacting surfaces.

In the selection of equipment for a process in an oil refinery the question of how much of it shall be electrical first must be considered. This will largely be governed by local conditions. Steam and electrical generating capacity of the plant have a great bearing on whether the larger power users such as compressors, pumps, etc., shall be motorized or not.

Of course in the larger refineries there is usually an ample supply of either of these utilities and the question then evolves itself into one of steam balance. Most of these refineries have combined steam- and power-generating plants where all or a great portion of the steam generated is at high enough pressures to pass through the turbines for the generation of power with the exhaust still retaining enough pressure and heat to drive steam motived equipment or to be used in the process. In these plants then a close study must be made upon all new demands for power to determine if it shall be in the form of electrical power or steam or a combination of the two.

Should the refinery have such a large demand for process steam that some must by-pass the turbines through reducing valves then new equipment should be motorized. On the other hand should the steam demand of the process be such that the generation of the power load produces more steam than the process could use it would be well to consider all new equipment as steam-driven.

However, most problems are not so simple and the usual solution admits both types of power drives especially where spares are necessary.

For instance on a modern unit it was desirable to use more motors than turbines in order to balance the exhaust steam system. At the same time if the unit were wholly motorized a power failure would be disastrous.

It was designed wherever possible with motor drive with steam turbine spares. The chearging pumps which were the largest consumers of power, were divided into two pumps in parallel, one, the larger, carrying three quarters of the charge motor driven and the other normally carrying one quarter but capable of carrying one half of the charge, turbine-driven. In event of power failure the turbine-driven pump would automatically increase to its full capacity which was sufficient to prevent coking until the electrical service could be restored.

But steam balance is not the wholly deciding factor in determining what shall be motorized. Continuity of service, variable speeds, starting torques, as well as other questions rise to make the choice difficult.

Doubtless a motor may be obtained to fit any condition if one is not averse to paying the price. Ordinarily the constant-speed induction motor or synchronous motor would seem to be first choice of most engineers, as these entail no complications in the way of starting equipment besides providing good efficiencies and power factors as well as low maintenance costs.

The requirements of flow control are such on most process units that these cannot be met with an ordinary slip ring motor, unless it were designed with a great number of steps in the resistance to give fine speed ad-

justments. Then also the waste of power entailed in such a method is considerable when it is considered that many refinery units operate over 7000 hours per year.

Flows of smaller quantities can be controlled successfully by using a type BTA motor with a pilot motor operated from a flow control.

A new development not yet commercially available is a thyraton motor which is expected to overcome most of the present disadvantages of a variable speed AC motor.

Gear motors can be used satisfactorily for small drives. They are especially adapted where a wide range of speed is desired at times and where the set-up for that condition need not be changed often.

However, the best flow control of quantities of oil is obtained from steam-driven equipment either alone or in conjunction with a motor as outlined previously.

The type of motor such as open, fan-cooled, splash-proof, or explosion-proof, to use depends upon the kind of service in which it is to be placed or the character of the locality.

In every refinery there are portions which may be classed as hazardous, others as semi-hazardous and some as non-hazardous. In making a selection of electrical equipment the type of area in which it is to be located must be kept in mind.

Standard open motors and starting equipment may be safely used in a non-hazardous area yet due to corrosive gases the upkeep and outages may be excessive.

Where this is true it may be worth while to use plated contacts or better oil-immersed equipment.

Although standard open type motors are still used to a great extent in the oil industry yet it would seem only a question of time until their use is discontinued. This will occur partly because they are not as safe as the newer types and will gradually be out-lawed by safety rules, but more because of the additional maintenance and upkeep required of them.

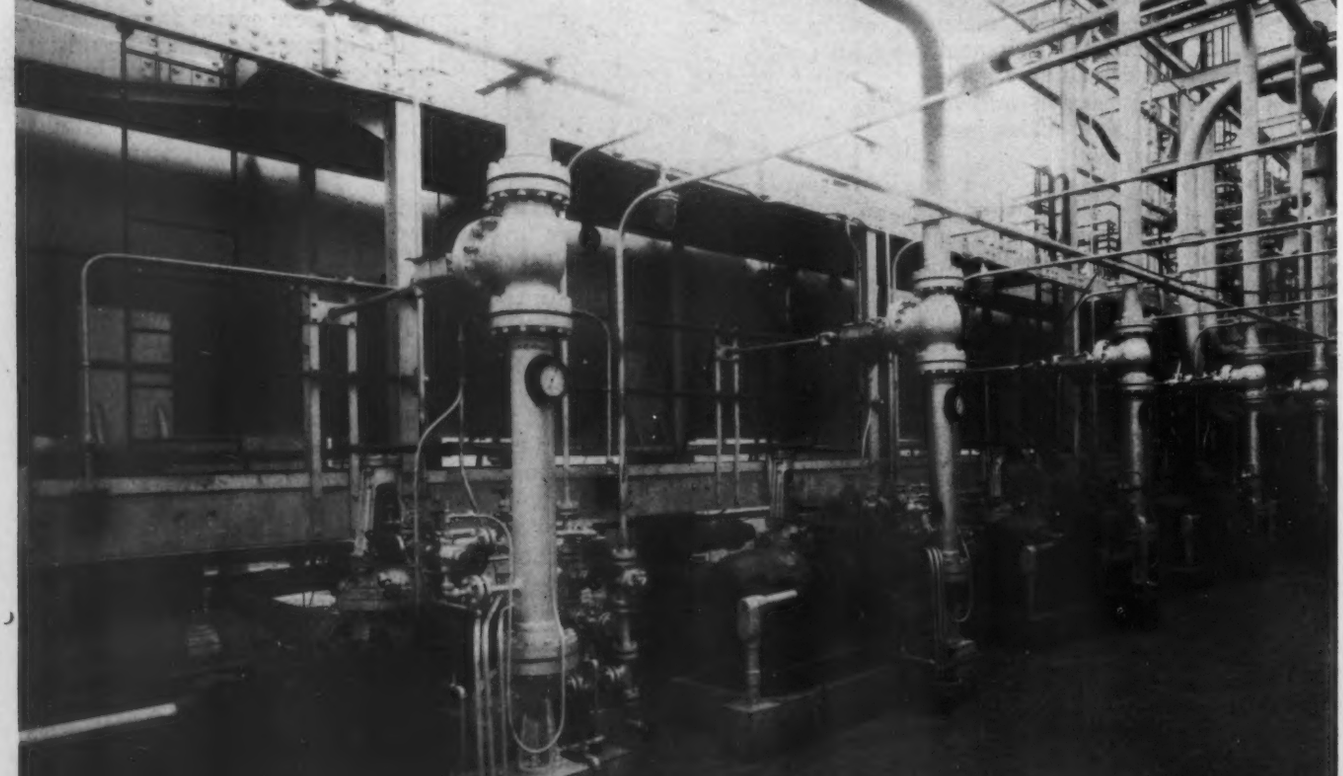
It is impossible to keep the windings of a standard motor used in an oil refinery free from oily dust and grease. This results in the necessity of numerous cleanings and eventually the breakdown of the insulation.

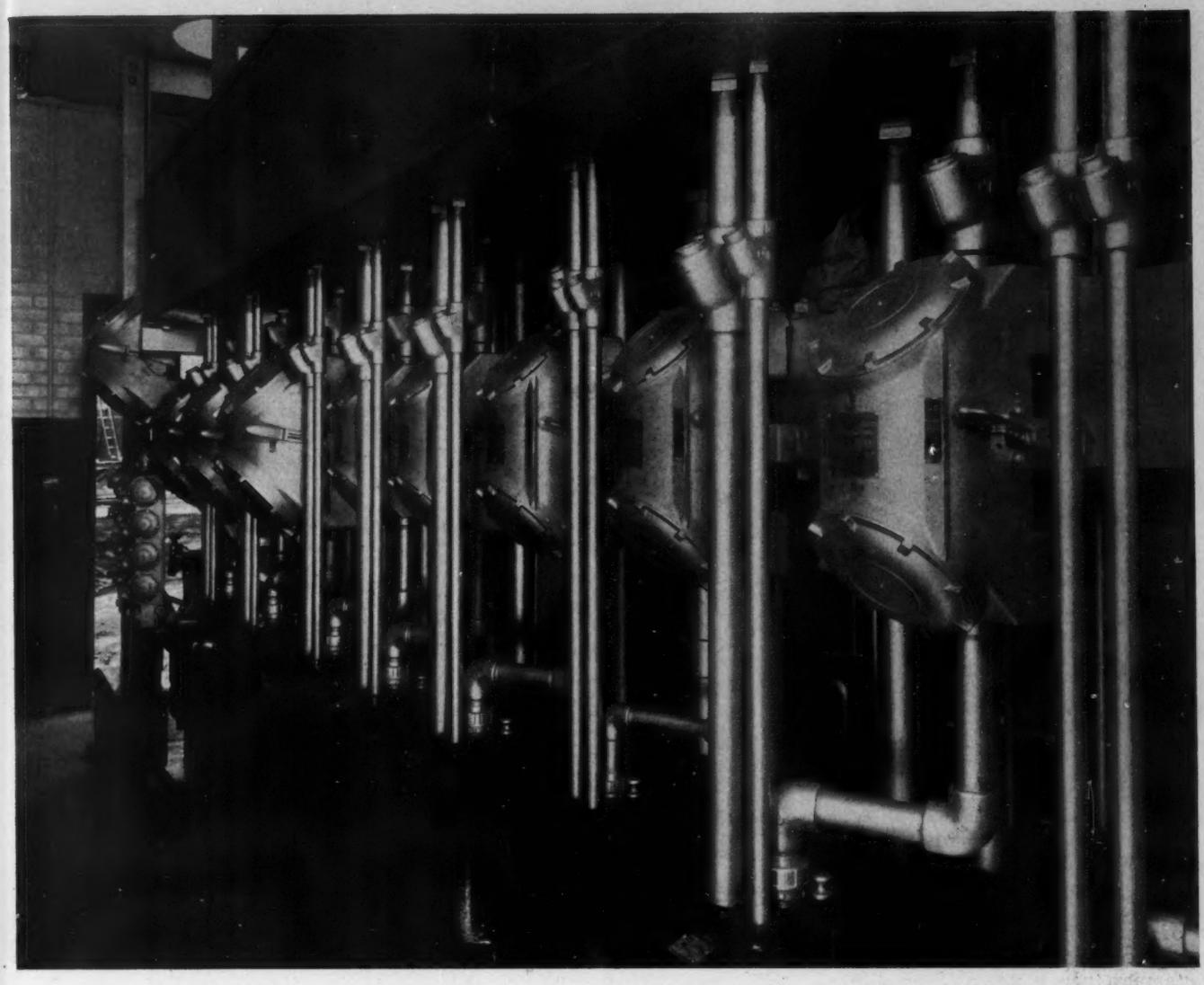
Totally-enclosed motors such as explosion-proof ones tend to prevent the destruction of the insulation from such causes and thus prolong the life of the motor, lessen the maintenance cost and reduce the number of outages of the equipment driven by them.

This type of motor is also useable for out-of-door installation without additional weather protection, a great advantage around refineries where the use of enclosures is avoided wherever possible from both the safety and economic standpoint.

The use of explosion-proof motors is becoming more and more general because nearly all process units have some areas that can be considered hazardous. Just how far such an area extends is dependent largely upon the local conditions existing at the moment. These may vary from day to day or even from hour to hour. Accordingly to be safe it is well to use explosion-proof motors

Explosion-proof motors, each coupled to a centrifugal pump operating out of doors in a large Texas refinery





Control room at the solvent refining plant in a large refinery. Lighting panel in the background.

even though they might not be installed in an area that could be termed hazardous.

In defining the extent of hazardous areas it is impossible to set any but arbitrary limitation. One state has formulated an electrical code for the oil industry within its borders and in it has resorted to arbitrary measurements in an attempt to define the limits of hazardous localities. A committee on motors, and motor controls which is a sub-committee of the Standardization Committee of the American Petroleum Institute is also up against the same problem in trying to set up rules or suggestions for the use of certain types of controls. In so far as motors are concerned it would seem well then to use only explosion-proof ones in those sizes in which they can be obtained. In the larger sizes and higher voltages which have not received underwriters approval it would be wise to use motors built along the same design as the smaller approved ones. An exception to this might well be made in the very large sizes where such a motor built to that design would be very expensive compared to a base ventilated one. If the ducts to the latter could be piped from a locality into which the chance of the introduction of explosive vapors would be remote then there should be no objection to its use even though the motor itself were installed in an area that might be termed hazardous.

In the selection of a motor it is sometimes necessary to take into consideration that at times process units are pushed to their limits. If there is a possibility that change of stocks will produce an additional load or if the average load is such that a motor to fit it must at times be called upon for additional capacity then the motor chosen should have a service factor great enough to carry this increase without damage to it.

With a standard open type 40°C. rise motor adequate overload protection is afforded when the temperature overload relay is rated at 115 percent of the rated load current of the motor. This gives a 15 percent "service factor."

Totally-enclosed fan-cooled motors of the Class I Group D construction do not normally bear a service factor on the same plate. The demand for enclosed motors in the popular sizes with the same dimensions as open motors, for a given horsepower rating, make it necessary for motor designers to use the highest temperature rise permissible for the type of insulation used. Since Class A insulation generally is used, a limiting temperature of 55°C. is applied.

If a motor were designed to actually rise 55 degrees, and if the ambient temperature were 40°C., as usually is assumed, no overload could be carried. As a matter

Explosion-proof disconnects on motor branch feeders, mounted out of doors.

of fact, however, the designers retain some margin under the limit of 55 degrees for often the ambient temperature is not as high as 40°C. This means that under such conditions these motors do have a small overload capacity. However, if an overload capacity as much as 15 percent is required, a larger frame with a normal load rating of 40°C. rise is necessary.

Frequently in the southwest where a great many refineries are located and where the normal summer temperatures are high the room in which a motor must be placed has at certain periods of the year high ambient temperatures. With an explosion-proof motor using Class A insulation a motor of size larger than actually necessary must be used to prevent excess temperatures. By using Class B insulation a motor of the size desired can be used with a gain in efficiency and price because with such insulation the motor can safely withstand the high temperatures.

This small gain in efficiency is very important when one considers that in the oil industry the time of operation is nearly continuous.

The voltages to be used on any installation are more or less standardized; 440 or 220 for the smaller motors with a decided preference for the former and 2300 volts for the larger ones.

The size of motor to be used on either voltage is limited somewhat by the manufacturers in that the smaller size are not standard in 2300 volts, but more by economic reasons in that it is cheaper to install the larger sizes at the higher voltages. Around 50 horsepower is a good dividing point although if only one or more motors in a group were but slightly larger or smaller than that it might be advisable to make them the same voltage as the rest of the group.

Across-the-line starting is the simplest form of equipment for energizing a motor and is satisfactory for most conditions. However, if a refinery is tied in with a utility in nearly all instances that companies rules apply to the size of motor allowed to be thrown directly across the line. Slow-speed motors present no difficulties in this respect even if of several hundred horsepower but motors of that size in the 3600 revolutions per minute speed class set up severe electrical disturbances on the line due to the greater inrush of starting current. These may be started with reduced voltage starters with considerable reduction in line surges.

Although starters and disconnects of the explosion-proof type may be installed at the motor yet the trend as recommended in the National Electrical Code is to install these in a control house situated in a gas-free area. This is a very desirable procedure as repairs which become necessary from time to time as well as frequent inspections, may be made in safety in the control house. In the event of repairs torches and soldering pots can be used there, an impossible act from the safety standpoint if attempted near the motor.

Occasionally it may be necessary to check the electrical load on the motors. Testing instruments may be used safely in such a control room. If these are inserted in the circuit near the motor a new hazard is created, nullifying the installed explosion-proof equipment.

Also the use of a control room in a gas-free area permits the use of starting devices and disconnects other than the explosion-proof types. This is imperative in the larger sizes. If one desires to use fused disconnects in place of more expensive ones of equal interrupting capacity it would then be permissible to do so. Whereas, in a hazardous area the use of fuses is limited to the small sizes in explosion-proof enclosures.

Until recently the fused knife switch was the only practical equipment available for a circuit disconnecting and short-circuit protective device.

There has been developed lately a very satisfactory enclosed and explosion-proof breaker for this service. It is manually operated, quick make and quick break, with an overload tripping device. It cannot be held closed against a short circuit or overload in excess of the value for which the tripping device is set.

With such a breaker, whenever the trouble in a circuit has been cleared service can be immediately restored without the time spent in obtaining and replacing the fuse. It has a feature valuable to a maintenance man in that the tripping element normally is sealed against tampering and can be changed only by substituting one of a different rating.

These breakers are able to interrupt currents of great magnitude because of their extremely rapid action in opening and their type of design.

Although the use of standard controls would be permitted in such a control room yet due to the corrosive vapors prevalent in a refinery the use of oil-immersed equipment, or better, oil-immersed explosion-proof types, will cut repair costs due to reducing maintenance due to corrosion to a minimum.

Motors with high inertia drives have a relatively long starting period. For this condition the temperature overload relay must be set so high that very little protection is given the motor. To meet this condition a saturated core current transformer is used between the relay and the line. By this means the current in the

heater element when starting is less than proportional to the line current. Proper selection will then permit starting with heavy currents flowing and the relay holding in, and still provide adequate protection on normal running loads.

Short-circuit protection can be provided on the smaller low-voltage motors by means of fuses if in a gas-free area or otherwise by explosion-proof air break or oil-immersed switches.

Thermal elements which are calibrated for one rating and cannot be changed without replacing the heater coil are very desirable in that they cannot be tampered with and do not allow setting beyond the safety range of the motor whenever a process desires to increase the load on the unit.

The higher voltage motors due to the larger interrupting capacity require that oil circuit breakers be used.

The question of interrupting capacity of breakers both on individual large motors as well as the main breakers of any unit is one that should never be overlooked, especially if there are possibilities of power growth with greater line and generating capacity. A great many refinery power systems generate at 2200 volts with no step-up or step-down transformers between the generators and the 2200-volt load. Unless current limiting reactors are used in each feeder circuit high interrupting capacity circuit breakers must be used. The cost to obtain breakers of sufficient interrupting capacity is negligible when compared to the expense incurred by the break-down of one of insufficient capacity.

Individual Circuit Breakers

Where a group of large motors have individual circuit breakers as well as a main breaker the individual breakers must have as great an interrupting capacity as the main if it is desired to lose only the motor on which a short circuit has occurred. If it is possible in the process to lose the whole group without serious trouble then circuit breakers of smaller interrupting capacity can be used provided they be latched in against such an eventuality with the main breaker acting.

Silver-plated contacts on disconnecting switches as well as on the contacts of circuit breakers due to the fact that silver oxides have nearly the same resistance as clean surfaces, tend to keep down temperatures. This is highly important on those circuits carrying heavy currents as inspections and cleanings need not be as frequent while the life of the oil in the switch boxes is increased due to the lower temperature and less carbonization.

Conduit systems should be ample for the installed electrical equipment with some spares in the main runs for future additional equipment. These cost very little when installed in the construction period but are sometimes very costly as well as unsightly if patched on the system later.

Conduits should be protected from mechanical dangers as well as fire hazards. Rigid conduit or fiber duct under ground encased in concrete gives adequate protection.

Precautions necessary for hazardous areas, such as sealing off conduits with sealing fittings are also good practice even when conditions are not considered hazardous. It is playing safe and at the same time keeping out corrosive vapors which might tend to increase maintenance.

Of course if an area is judged dangerous enough to warrant the use of explosion-proof motors common sense suggests on all new constructions the installation of explosion-proof lighting fixtures and a complete explosion-proof conduit job including explosion-proof receptacles and hand lamps if such be necessary.

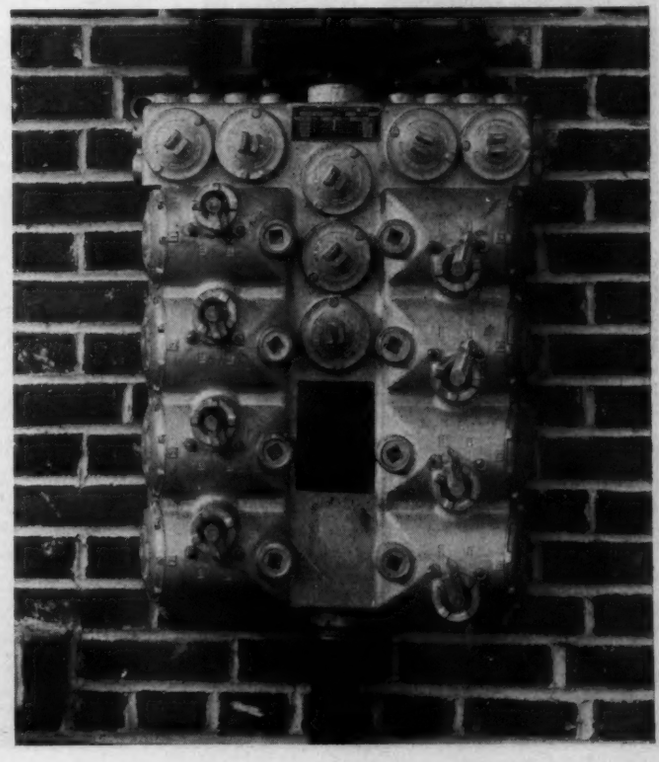
The use of explosion-proof type of condulets (fixtures and switches excepted) is not going to increase the first cost greatly while the benefits from the maintenance standpoint will soon pay the added cost. A completely sealed system means no trouble from corrosive gases entering switches while the condulets with screw covers offer greater convenience in making connections and in insuring a tight job.

Fully-enclosed lighting panels save the maintenance man much annoyance, especially where the panels used are explosion-proof with some form of breaker instead of a fuse.

Lighting panels of the standard type if exposed to corrosive gases require considerable attention even though switches, fuse clips, and fuse contacts are plated with some corrosion-resisting metal.

Conduits run in the open on units subject to numerous flashes or fires often cause excessive maintenance. One company, around the furnaces of a cracking unit where flashes are frequent, used a method of protection that has paid for itself many times over since its installation. All conduits of power, lighting, or thermocouple lead wire are covered with pipe covering similar to that used for steam lines. As additional protection this covering is encased with a thin metal sheet. Whenever possible the lines are grouped and one covering placed over the group. With this protection flashes burn nothing but the flexible connections to the couples and some of the smaller branches of power and lighting. It must be a severe fire to injure any of the main conduits. At the time this was first installed thermo-

Explosion-proof lighting panel mounted out of doors.



couple lead wire costing $0.30 per foot was being used. Consequently it paid off in a hurry, to say nothing of the time saved in putting the unit back in service.

Good looking open conduit runs do not necessarily decrease maintenance but they are the mark of good workmanship and certainly every worthwhile construction man strives for a good looking job. Where bends are necessary and standard elbows do not make a uniform job it is well with a number of conduits running parallel, to make a form or template for the entire group and fit all bends and offsets in it before assembling in the run. To do good work on medium and large size conduit, it is necessary to use some form of bending device in order not to crush or kink the conduit. Home made benders can be constructed at small expense that are serviceable and satisfactory.

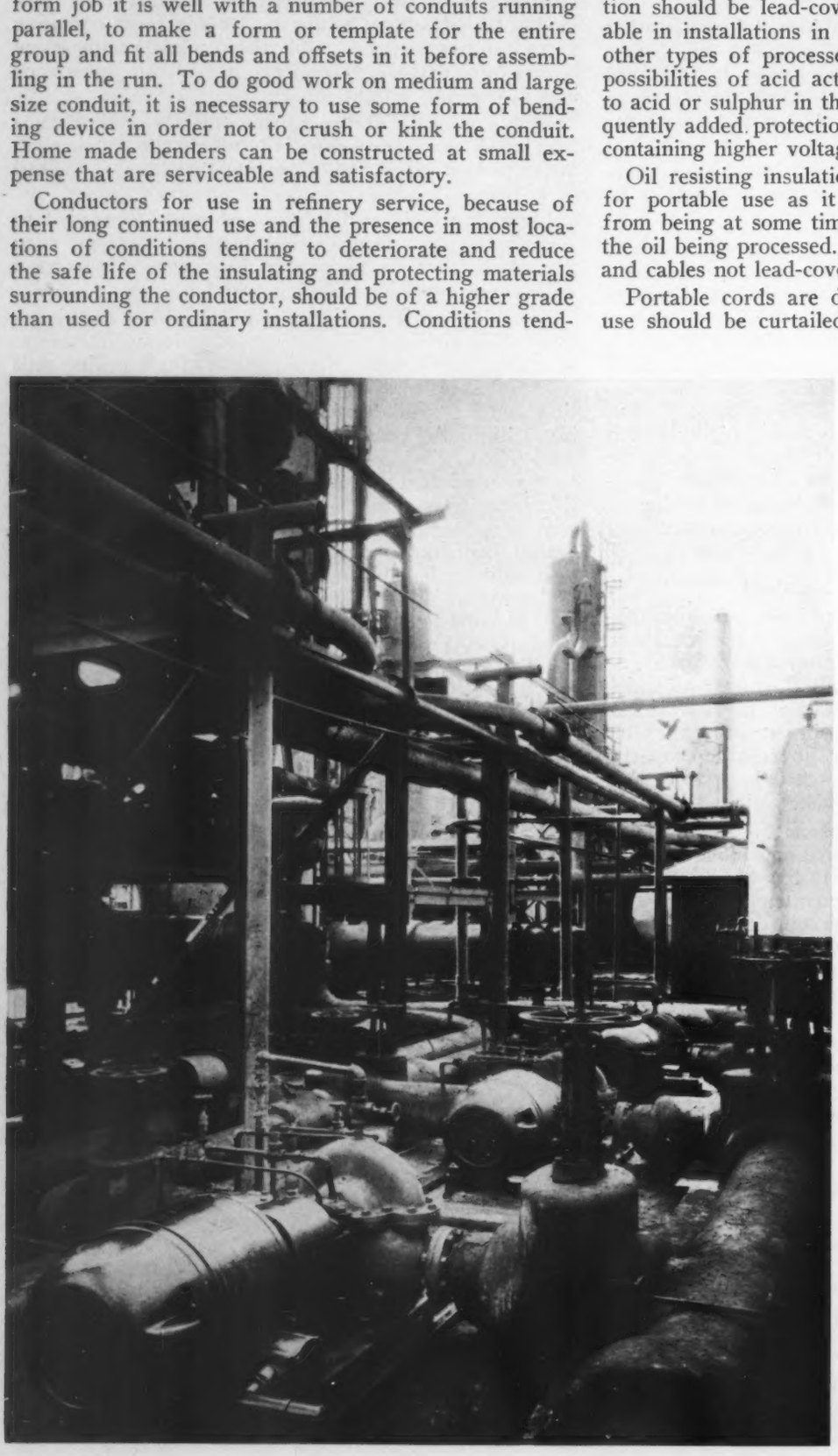
Conductors for use in refinery service, because of their long continued use and the presence in most locations of conditions tending to deteriorate and reduce the safe life of the insulating and protecting materials surrounding the conductor, should be of a higher grade than used for ordinary installations. Conditions tending to injure the installation may be long hours with moderate or higher temperatures, moisture, oil, chemical fumes, etc.

The use of stranded conductors on all but smallest sizes facilitates installation and removal and avoids breaking.

Underground cables where there are possibilities of moisture entering through leaks or through condensation should be lead-covered. This is particularly desirable in installations in older refineries where formerly other types of processes had once been installed. The possibilities of acid action of the soil on conduits due to acid or sulphur in the ground are very great. Consequently added protection on the cables, especially those containing higher voltages, may be of advantage.

Oil resisting insulation should be used on all cables for portable use as it is impossible to prevent them from being at some time in contact with some form of the oil being processed. In fact practically all the wires and cables not lead-covered should have this insulation.

Portable cords are dangerous implements and their use should be curtailed as much as possible. A good



Explosion-proof motors, coupled to centrifugal pump outdoors in refinery yard.

lighting system will minimize their use. However, when they must be used, approved devices should be installed and these should be thoroughly tested and inspected regularly.

Where volatile liquids or gases are handled in the process, grounding is a good protection against possibilities of static discharges. Of course static charges can and do occur where heavier oils are handled but the possibilities of ignition occurring are much less. In the dryer climates static charges build up more readily while in those of great humidity such as in the Gulf Coast region these do not seem to be so prevalent.

True maintenance consists of inspections and minor repairs made frequently enough to forestall major repairs until such become necessary by the natural course of time.

It is impossible for any repair man to be certain of time of inspection and condition of the apparatus unless records are kept. These may be very detailed or not, depending upon how closely one wishes to check minor repairs. They should, however, list dates of inspections and all major changes.

A card index listing all the essential data of each piece of apparatus to be inspected is of great value. A form shown is one issued by one of the larger electrical manufacturers in listing motors. Repair parts bulletins bound in book form then may be referred to readily from this data, making an invaluable combination in obtaining and replacing damaged parts. The ability to at once locate the manufacturers catalogue numbers of the parts needed is a source of satisfaction to anyone who has wrestled with the problem of maintenance.

If many changes or repairs are necessary it is helpful to keep a complete history. As this requires considerable space a loose-leaf file with individual sheets for each unit provides it.

Where a major repair is made, such as on a large turbine drive, then a complete record of all repairs and clearances must be kept for these details are invaluable at the next inspection in checking whether faults have been corrected. It should show the actual condition of the machine as found with respect to valves, governor, mechanism, bearings, wheel clearances, and if possible vibration readings before it was shut down. Any evidence of corrosion should be noted carefully. Where this is taking place the use of photographs at each internal inspection will show visually to what extent this is progressing and whether the steps taken to combat it have been successful. When the machine is reassembled records should be made of wheel clearances and vibrations.

Where a main control room is used with numerous conduits leading to the various parts of the unit, restoration of service after interruptions is speeded by the proper labeling of the various controls either by name or number corresponding to the apparatus driven. The danger of tripping out the wrong circuit whenever any apparatus is to be shut down is also avoided.

As maintenance men often are changed or regular men not available during an emergency it is a great aid to have at hand a print of the various circuits and the apparatus connected thereto. This print can be framed and hung on the control room wall, or if too large may be kept rolled in a suitable metal container.

Whenever an inspection reveals some repairs are necessary beyond the minor ones that the inspector can perform readily, some form of report in writing should be made. This should go to the one who has the proper authority to investigate further and issue a work order should it be deemed necessary to make these repairs. Again, how detailed this report should be or through what hands it should pass depends upon the set-up within the company or department.

On the larger refining units it is impractical to make more than minor repairs on the main charging drives or fan motors while the unit is in service. Consequently all work orders are saved until the process unit comes "off stream" when an intensive repair program takes place. At that time rigid inspection must be made that no potential fault be overlooked and bring the unit down off schedule.

Even the routine of oiling and greasing bearings should be made a matter of record with these kept for reference by some one in authority so that the mechanic's acts may be checked in case of failures of bearings either to lack of attention or to the improper use of oils and greases.

An aid in forestalling trouble is some form of ground detector. This of course would not be explosion-proof, but where a control room is used an instrument or a bank of lamps will readily show that such trouble has occurred and means can then be taken to correct it before any further damage can take place. Meggar readings taken and recorded regularly are also a great aid in guarding against a gradual break down of insulation.

H. P.	Syn. Speed	Make				Winding		Shop or Mill No.	
3	1800	General Elect. Co.						B.B.C.S 322	
Type	Frame	Poles	F.L. Speed	Volts	Form	Temp. rating	F.L. amp.	Phase	Cycles
KT	932	4	1780	220	C	40°C	7.76	3	60
Serial No.		Mfr's Order No.		Our Order No.		Date of Order		Cost	
4926092									
Model No.		COILS		CONTROL EQUIPMENT					
63A2		Stator or Field Spec.		Kind Oil circuit breaker					
BRUSHES		Main		Make type FP-110-T-P.S.T. 50A 600V					
No.		Comm.		Type R-P. 2710-2602727-G.S					
Cat. No.		Comp.		Serial No. overload coil 8A.					
Size		Rotor or Arm. Spec.		Speed Range					
Material		PULLEY		Accessories					
Grade		Diam.							
BEARINGS		Face							
Front End Cat. No.		Bore							
Pulley End Cat. No.		Keyway							
REMARKS:									

A Simplified Method for

Absorber Design

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Humble Oil & Refining Company
Baytown, Texas

ABSORPTION of the soluble components from a vapor mixture by means of a liquid solvent in a continuous countercurrent absorber has been made the subject of numerous studies, and the physical laws which govern the process are now well known. In general, the equilibrium relations of the one or more soluble components, the material balances on each component throughout the absorber, and the number of equivalent theoretical plates determines the composition of the streams from the tower. In the graphical solution of an absorption problem, an equilibrium curve is constructed from vapor pressure data, an operating line is constructed from an equation representing a material balance, and the number of theoretical plates required is found by counting the steps in a "staircase" constructed between the equilibrium and the operating lines. It is frequently desired to calculate the performance of absorbers recovering casinghead from a complex mixture of field or refinery gases, and particularly to estimate the amount of butane recoverable at various operating conditions. In solving a problem of this type the calculations can become very laborious, due partly to the large number of components present in the wet gas entering the absorber; moreover, the data obtained on operating equipment must be supplemented by various assumptions, and during the calculations the effect of these assumptions on the final result can be easily obscured. In order to overcome these difficulties, a systematic method of performing the work has been developed and is presented here. The new features incorporated in this method are first, a choice of coordinates such that the operating line is always linear with a slope of unity, and second, the use of equations in which the units of measurement customarily employed in the plant can be substituted directly.

In the absorption of refinery gases the operating conditions, as determined by economic considerations, are usually such that the recovery of pentane and of all components of higher boiling point is substantially complete; moreover, the rich oil is saturated with respect to propane and all components of lower boiling point, so that the operation and design of absorbers are usually controlled by the recovery of the butane fraction. Even when there is sufficient variation in the recovery of the other components to be of economic importance, the butane recovery should still be considered because of its effect on the molar overflow and tower temperature (due to heat of solution). For this reason the method has been developed particularly with regard to butane.

A basis of 1000 cubic feet (1M cu. ft.) of wet gas, measured at 60°F. and at atmospheric pressure, has been chosen. A table of nomenclature defines the symbols used below. The operating line for butane is given by the equation:

$$B_v = Bo + B_D \qquad (1)$$

which states that the M cu. ft. of butane rising to any plate is equal to the M cu. ft. of butane dissolved in the overflow from that plate plus the M cu. ft. of butane leaving in the dry gas.

The equation for the equilibrium line is:

$$B^* = k\left(\frac{P}{\pi}\right)\frac{Bo}{(0.379\,O + V_{KO})} \qquad (2)$$

where Bo, the butane dissolved in the overflow, is the independent variable. While this equation is not theoretically exact, it has been found that further refinements do not appreciably change the position of the equilibrium curve. A Raoult's Law correction factor, k, is shown; a value of 1.1 is recommended. The following procedure is given for evaluating the unknowns in the above equation.

CALCULATION OF THE KNOCKOUT V_{KO}

When the gas streams to and from the absorber are metered, the knockout is obtained by dividing the contraction in volume by the volume of wet gas. When one

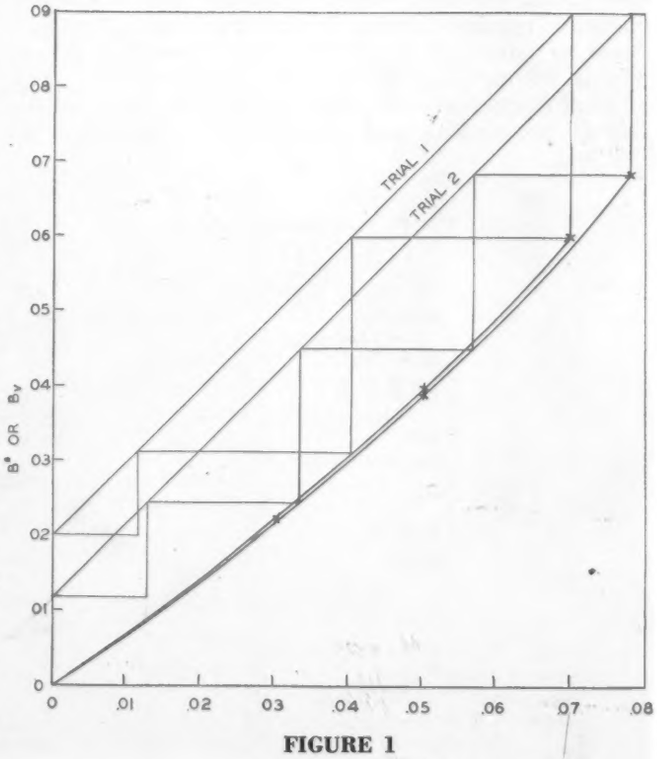

FIGURE 1

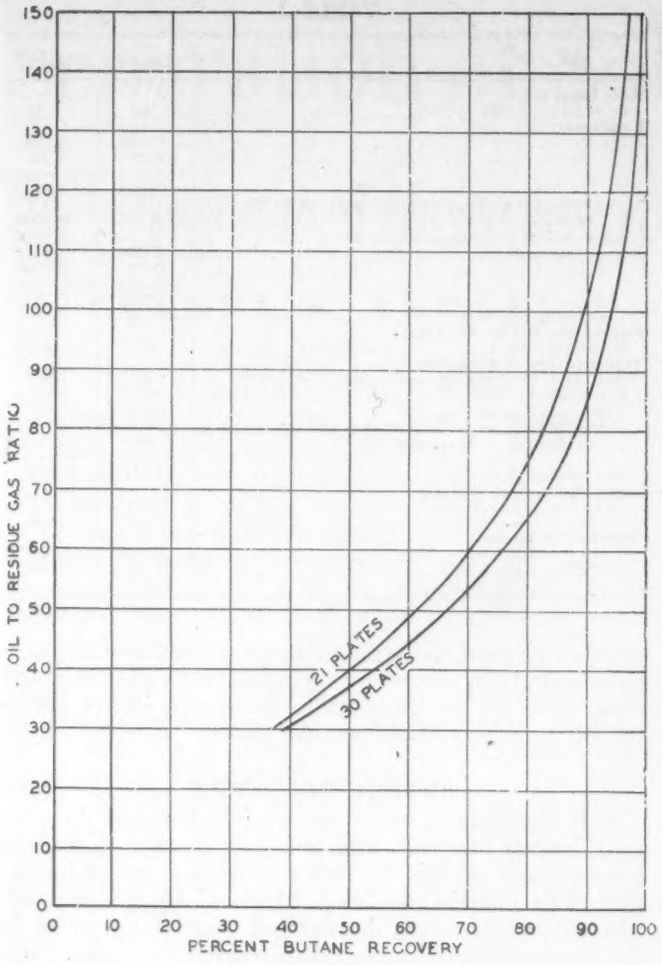

FIGURE 2

of the gas streams is unmetered, but the temperatures of all the streams are known, the knockout can be calculated by the use of Equation (5) below. In the absence of test data or in the design of absorbers the knockout can be calculated from the oil-gas ratio and the composition of the wet gas by either of the following equations, the first to be used when the oil-gas ratio is expressed as mols of lean oil per M cu. ft. of wet gas, and the second when the oil-gas ratio is expressed as mols of lean oil per M cu. ft. of dry gas.

$$V_{KO} = Y_s + B_{KO} + 0.5 \frac{\pi 0}{P_s} (Y_s + 0.1\ Y_L) \qquad (3)$$

$$V_{KO} = \frac{Y_s + B_{KO} + 0.5 \dfrac{\pi 0'}{P_s} (Y_s + 0.1\ Y_L)}{1 + 0.5 \dfrac{\pi 0'}{P_s} (Y_s + 0.1\ Y_L)} \qquad (4)$$

The above equations assume complete recovery of pentane and heavier. The propane and lighter dissolved is given by an empirical term which has been found to check satisfactorily with plant tests. It will be noted that this term shows the amount of propane and lighter to be inversely proportional to the vapor pressure of propane at the lean oil temperature. The amount of absorbed butane, B_{KO}, can be roughly estimated by multiplying the mol fraction of butane in the wet gas, Bw, by the percent butane recovery read from Figure 2. At this stage a rough approximation of B_{KO} is satisfactory, as the resulting error in the knockout will be too small to alter noticeably the position of the equilibrium line which is to be used in the plate-to-plate calculation of B_{KO}. The value found can then be substituted into the equation if it is desired to obtain the correct value for V_{KO}, the total knockout. If the dry gas is metered, but not the wet gas, the wet gas rate can now be obtained by dividing the dry gas rate by $(1 - V_{KO})$.

CALCULATIONS OF MOLS LEAN OIL

Usually the lean oil rate is known in gallons. In order to convert this to mols, it is necessary to multiply the lean oil rate by the lbs. per gallon corresponding to the gravity of the lean oil, and to divide by the molecular weight. The molecular weight of the lean oil is equal to the molecular weight of a pure paraffin hydrocarbon having the same boiling point as the temperature at 50 percent off on the A.S.T.M. distillation curve. Mols of lean oil is divided by M cubic feet of wet gas in order to obtain O, used in Equation (2).

CALCULATION OF THE RICH OIL TEMPERATURE

In the absence of test data, or in design problems, it is necessary to estimate the temperature of the rich oil in order to find the vapor pressure of butane at the bottom of the tower. The rich oil temperature is higher than the lean oil temperature due to heat of solution of the dissolved components. The temperature of the rich oil can be calculated by a heat balance, which is conveniently expressed by the following equation:

$$T_{\text{rich oil}} = T_{\text{lean oil}} + \frac{C\,V_{KO}}{R} - 8\frac{T_{\text{dry gas}} - T_{\text{wet gas}}}{R} \qquad (5)$$

The heats of solution of various pure paraffin hydrocarbons are found to be a linear function of molecular weight, from which:

$$C = 120\ (\text{Ave. MW of Knockout}) - 400 \qquad (6)$$

The calculation of the rich oil temperature by this method has been checked by several plant tests.

It is assumed that P, the vapor pressure of butane, varies linearly with Bo. It is also assumed that the average vapor pressure of the butane fraction will be somewhat nearer to that of normal butane than to that of isobutane. The tower pressure, π, is expressed in the same absolute pressure units as P.

NUMBER OF PERFECT PLATES

It has been shown how each unknown of Equation (2) can be calculated. Of these unknows, all are constants for a given operation except Bo, the independent variable, and P, the vapor pressure of butane, which is a function of Bo. The equilibrium value of butane, B*, can then be calculated for several values of Bo, and the equilibrium curve, B* vs. Bo, is then constructed. The operating line is also laid off with B_v plotted on the same scale as B*, both as functions of Bo. The slope of the operating line is unity, and the intercept is B_D, the M cu. ft. of butane in the dry gas per M cu. ft. wet gas. A "staircase" is constructed between the equilibrium and the operating lines, and the number of perfect plates is counted. The number of perfect plates divided by the number of actual plates is the plate efficiency, discussed below. If the number of perfect plates is known, and B_D is unknown, a value of

B_D is found by trial and error which will check the known number of perfect plates.

ILLUSTRATION OF METHOD

A typical absorption problem is presented to illustrate the method. It is desired to calculate the percent butane in the residue gas from the absorber, and the corresponding butane recovery, from the following operating data.

Composition of Wet Gas:

Methane	40.0 percent
Ethane	25.0 percent
Propane	22.0 percent
Butane	9.0 percent
Pentane	4.0 percent
	100.0 percent

Actual number of plates = 21
Plate efficiency = 18 percent
Oil to residue gas ratio = 90 gals./M cf.
Molecular weight of lean oil = 220
Density of lean oil = 7.3 lbs. per gallon.
Temperature of lean oil entering tower = 90°F.
Temperature of rich gas entering tower = 90°F.
Temperature of dry gas leaving tower = 95°F.
Tower pressure = 63 lbs. per sq. in. ga. = 5.3 atm. absolute.

It is seen that the equilibrium curves from the two trials are for practical purposes identical; one trial was therefore sufficient to obtain the equilibrium curve, the first trial being shown merely for comparison. The corresponding operating lines are shown on Figure 1. In the first trial 2.4 perfect plates are required, whereas 3.8 perfect plates are available (21 actual plates, 18 percent efficiency). In the second trial the operating line was shifted (by trial and error) to give 3.8 perfect plates. It is not at all difficult to locate the proper position for the operating line if it is remembered that the operating line must average 0.078/3.8 = 0.021 units to the left of the equilibrium curve.

From Figure 2 it is found that 0.012 M cubic feet of butane per M cubic feet of wet gas remain in the residue gas. The percent butane in the residue gas is found by dividing by $(1 - V_{KO})$:

$(0.012/0.73)\ 100 = 1.6$ percent butane in residue gas.

The butane recovery is found by dividing the M cubic feet of butane dissolved in the rich oil by the M cubic feet of butane in the wet gas, or:

$(0.078/0.09) = 87$ percent butane recovery.

PLATE EFFICIENCY

A number of tests have been made on absorbers in order to estimate their plate efficiencies. Calculations, made by means of the above method, give an average plate efficiency of 18 percent over wide variations in the operating conditions. A typical absorber consists of a six-foot diameter tower with 21 plates on 20-inch centers. The bell caps are 6½-inch diameter, with 32 slots, ¼-inch by one-inch, with about 50 bell caps per plate. The bell caps and overflows are maintained in reasonably good condition. The same efficiency has also been found for absorbers equipped with 4½-inch bell caps. The top of the slots is about two inches below the oil level maintained on the plates.

TABLE 2

Trial I:	Rich Oil	Intermediate Points		Lean Oil
B_O	0.07	0.05	0.03	0.0
P	4.55	4.2	3.85	3.3
$B^* = 1.1 \frac{(P)\ B_O}{(5.3)\ 1.10}$	0.06	0.04	0.022	0.0
Trial II:				
B_O	0.078	0.05	0.03	0.0
P	4.6	4.15	3.8	3.3
$B^* = 1.1 \frac{(P)\ B_O}{(5.3)\ 1.10}$	0.068	0.039	0.022	0.0

TABLE 1

	Trial I	Trial II
Mols Lean Oil:		
$O' = 90 \times 7.3/220$	3.0	3.0
Knockout:		
$0.1\ Y_L$	0.065	0.065
Y_3	0.22	0.22
	0.285	0.285
P_3 (vapor pressure propane, atm. @ 90 °F)	11	11
$0.5\ \pi\ 0'\ (Y_3 + 0.1\ Y_L) / P_3$	0.205	0.205
$B_{KO} = 0.09 \times 0.87$ (from Figure 2)	....	0.078
B_{KO} (Assumed)	0.07	....
Y_5	0.04	0.04
	0.315	0.323
V_{KO} (Using Equation 4)	0.26	0.27
Ratio Lean Oil to Wet Gas:		
$R = 90\ (1 - V_{KO})$	67	66
Temperature of Rich Oil:		
Ave. Mol. Wt. of Knockout	50	50
$C = 120 \times 50 - 400$	5600	5600
Temperature Lean Oil	90 °F.	90 °F.
$C\ V_{FO}/R$	21.8	22.9
-8 (T dry gas — T wet gas) /R	—0.6	—0.6
Temperature Rich Oil	111 °F.	112 °F.
Vapor Pressure of Butane:		
At lean oil temperature	3.3	3.3
At rich oil temperature	4.5	4.6
Overflow in Tower:		
$0.379\ O = 0.379\ O'(1 - V_{KO})$	0.84	0.83
$0.379\ O + V_{KO}$	1.10	1.10

BUTANE RECOVERY

It is desirable to know the relation between the percent butane recovery and the oil-gas ratio for plant control purposes and for cost studies of absorber operation. This relation has been calculated for 21- and 30-plate absorbers, using the wet gas composition given in the above illustration, for oil to residue gas ratios from 30 to 150 gallons per M cubic feet of gas. It is suggested that these curves (Figure 2) be used as a first approximation for the value of B_{KO} in preliminary trials.

The method herein presented lends itself readily to the solution of numerous problems which have gone unsolved in the past due to the laborious computation required. With a minimum of effort it should now be possible to determine the effect of varying the composition of the wet gas, the lean oil temperature, the tower pressure, etc., on the recovery of gasoline fractions in the absorber.

NOMENCLATURE

B_D = M cu. ft. of butane in dry gas per M cu. ft. wet gas.
B_V = M cu. ft. of butane in vapor to plate per M cu. ft. wet gas.
B_O = M cu. ft. of butane dissolved in overflow from plate per M cu. ft. wet gas.
B_{KO} = M cu. ft. of butane dissolved in rich oil per M cu. ft wet gas.
B^* = M cu. ft. of butane in equilibrium with overflow from plate per M cu. ft. wet gas.
V_{KO} = Total M cu. ft. dissolved in Rich Oil per M cu. ft. wet gas.
R = Gallons lean oil per M cu. ft. wet gas.
O = Mols lean oil per M cu. ft. wet gas.
O′ = Mols lean oil per M cu. ft. dry gas.
Y_5 = Volume percent pentane in wet gas.
Y_3 = Volume percent propane in wet gas.
Y_L = Volume percent lighter than propane in wet gas.
π = Tower Pressure.
P = Vapor pressure of butane fraction.
P_3 = Vapor pressure of propane at lean oil temperature.
K = Correction factor to Raoults law = $\pi y/PX$.
C = Function of M. W. of knockout, used to calculate the temperature of the rich oil.

Separation of Solvents and Water *in Refining and Dewaxing of* Lubricating Oils

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and
WHEATON W. KRAFT
The Lummus Company, New York

THERE has been rapid development in processes employing solvents for the production of high quality lubricating oils from crude petroleum. These processes include solvent refining and solvent dewaxing. The solvents used commercially include acetone, aniline, benzol, chlorex ($\beta\beta'$ dichloro diethyl ether), cresylic acid, carbon tetrachloride, ethylene dichloride, furfural, methyl-ethyl ketone, nitrobenzene, phenol, propane, liquid sulphur dioxide and toluol. The characteristics of these solvents have been analyzed and their effect on the choice of the process to be employed for solvent-water separation in solvent refining and dewaxing plants has been indicated. (Bibliography of solvent properties has been prepared). Included in this paper are flow sheets of typical solvent refining and solvent dewaxing processes. The importance of separating solvent and water in these plants has been indicated. The various methods for separating water and solvent have been discussed and a number of flow sheets illustrating these methods are included. In general, fractional distillation appears to be the most suitable method because of its adaptability to solvent recovery systems and the familiarity of the refiner with this type of equipment. It is necessary to study each problem and then decide which process is most suitable.

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DURING the past few years there has been rapid development in processes employing solvents for the production of high quality lubricating oils from crude petroleum. This development has been such that it is now possible to produce lubricating oils of good marketable quality from crude oil formerly considered to be unsuitable for lubricating oil production. The more rigid specifications for lubricants which have resulted from the development of the automobile engine to its present state and the widening temperature range over which industrial lubricants are being used has made necessary the commercial use of processes capable of producing lubricants to meet these specifications. The value of solvent processes for production of high quality lubricants is obvious. Of more general importance, however, is the fact that the decreased cost of such related processing steps as decolorization, clay treating, etc., may lead to substantial operating economies.

Solvent processes have for their main object the removal of such material as tars, resins, waxes, and various related groups. The presence of this material leads to undesirable carbon formation, rapid oxidation and sludging, high solidification temperature and rapid change in viscosity with temperature. Solvents used commercially include acetone, aniline, benzol, Chlorex (β, β'—dichloroethyl ether), cresylic acid, carbon tetrachloride, ethylene dichloride, furfural, methyl ethyl ketone, nitrobenzene, phenol, liquid propane, liquid sulphur dioxide, toluol, and trichloroethylene. Individual solvents or mixtures of two or more solvents are used. Where a solvent mixture is used, the composition of the mixture is controlled to obtain the definite characteristics desired. Where more than one solvent is used, the application of each solvent may be in successive steps, or simultaneously. The latter procedure is followed in one commercial process in which two solvents of opposite characteristics are used.

IMPORTANCE OF SEPARATING SOLVENT AND WATER

Since water is practically always present in commercial operation of solvent plants, even in small concentrations, its separation from the solvent is of especial importance for two reasons: first, most solvents must have a controlled water concentration if the characteristics inherent in a desirable selective solvent are to be maintained; and second, any solvent associated with excess water removed from the system represents a direct operating loss. The economic importance of the second point is evident from a consideration of the cost of solvents, and of the magnitude of solvent losses which would result if no provision for recovery were made. The presence of water in the system can be readily accounted for. It may enter the system with the oil or solvent charge or through leaks in water cooled condensing and cooling equipment. When a plant is shut down for periodic inspection, repairs or other causes, the entire system is normally drained free from liquids and steamed out to completely eliminate all material of either a toxic or inflammable nature. This operation obviously results in the accumulation of considerable water in the system. When steam is used in the process for stripping the final traces of solvent from the oil, extract or wax products, the water introduced will be appreciable.

OCCURRENCE OF WATER IN SOLVENT PLANTS

A brief outline of solvent refining and dewaxing processes will show the points where water appears.

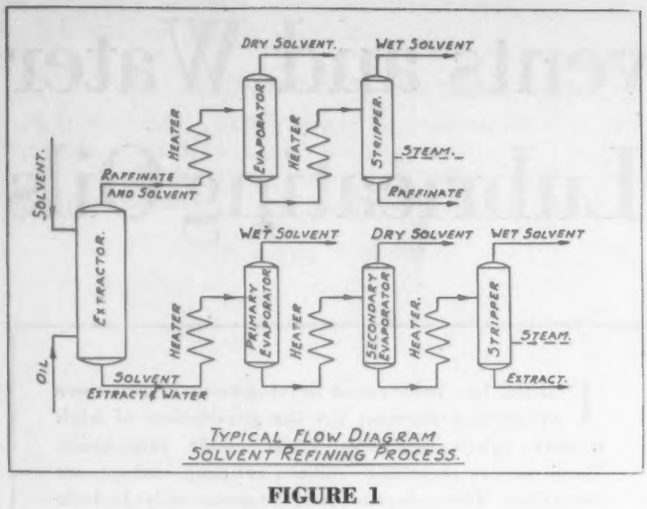


FIGURE 1

Figure 1 illustrates diagrammatically the essential steps in solvent refining. The raw oil to be processed is extracted by the solvent with subsequent phase separation. The resulting solvent phase contains the extracted material (tars, resins, etc.) and the oil phase contains refined oil (raffinate) with a varying percentage of dissolved solvent. In general, water present in the system at this point will concentrate in the solvent phase. Each phase is then separately distilled in suitable apparatus for recovery of solvent, the recovered solvent being re-used in the process while the oil and extract continue to other processing equipment. Most of the solvent can be recovered as water-free solvent, but a certain percentage is associated with water. The disposition of this latter material constitutes the problem to be solved.

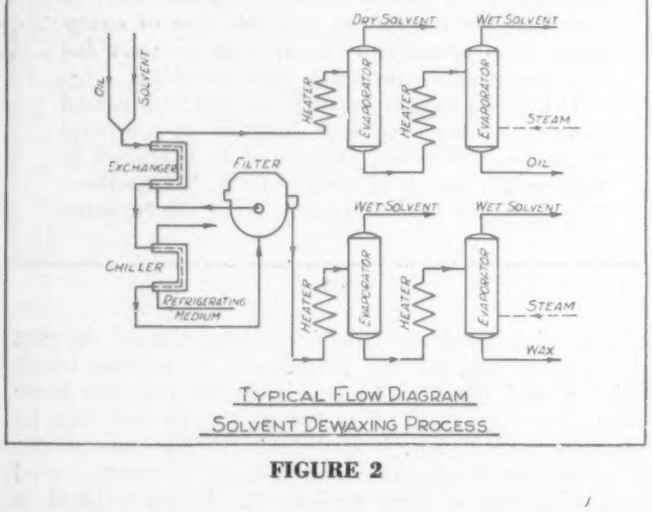


FIGURE 2

Figure 2 illustrates diagrammatically the essential steps in solvent dewaxing. The wax-bearing oil is dissolved in a suitable solvent or solvent mixture, and the solution subjected to controlled chilling. Wax is thrown out of solution by this step and is separated from the solution by centrifuging or filtering. Any water in the mixture subjected to chilling simply solidifies. If the separated wax is centrifuged, the ice is associated with wax rather than oil solution because of its specific gravity. If the separated wax is filtered, the ice becomes a part of the wax cake formed on the filters. Water in the system thus is associated with the separated wax. Solvent is recovered from the oil solution and wax by distillation, and is returned to the process for re-use. The equipment for recovery is of the same general type as that used in solvent refining. Flow diagrams of the more widely used solvent processes for refining and dewaxing have appeared in various publications.[1, 2, 3, 4, 5, 6, 9, 11, 12, 13, 15, 16, 17, 18, 19]

ELIMINATION OF WATER SOURCES

The reasons for the presence of water in a solvent plant have been previously explained. Obviously, introduction of water from the sources mentioned can be minimized or entirely prevented. If the charge oil contains much water, it may be dehydrated before being delivered to the system for processing. There are a number of methods for removal of the last traces of solvent from oil without the use of stripping steam. Heating of the oil under a high dry vacuum is one method. Removal of the solvent by chemical treatment is another. One plant using an organic acid, effects removal of the final traces of the solvent by treatment with a solution of caustic soda. After separation from the oil, the resulting salt solution is then treated with strong acid to set free the organic acid. The complications of this method are obvious. An interesting method is that proposed by D. E. Stines[10]. This is shown diagrammatically in Figure 3. Heated oil-solvent mixture is

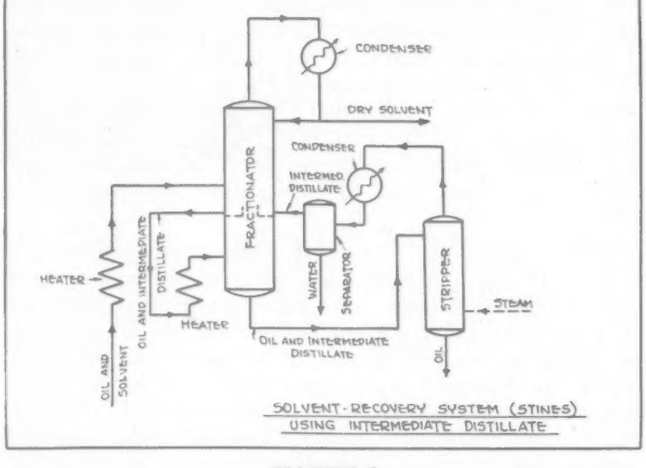


FIGURE 3

introduced at an intermediate point of a fractionating column. In the lower section of this column, material added prior to the starting up of the operation accumulates. This material has a boiling range intermediate between that of the solvent and that of the oil. Near the bottom of the column a stream is withdrawn, circulated through a heater and discharged into the bottom chamber of the column. A temperature high enough to vaporize this intermediate material is maintained so that the vapors of intermediate material rising from the bottom chamber strip solvent from the liquid oil-solvent mixture flowing down the column. In so doing, the vapors themselves are condensed and flow back down the column with the now solvent-free oil. By this system the mixture subjected to heating contains no solvent, which is important if the solvent used is easily decomposed. The bottoms product from the column still contains some of the intermediate material, but this is removed by steam stripping in a separate stripping column provided with a condenser. A narrow boiling range

petroleum fraction is suggested as the intermediate material, since condensation of the mixed vapors from the stripping column then provides an easily separated liquid mixture. The water is decanted and the petroleum fraction returned to the main column. Among other suggested methods may be mentioned one in which an inert gas is used as a stripping medium in place of steam and another in which propane vapor is similarly employed. The second method is especially applicable where liquid propane is one of the solvents used in the process.

PROBLEM OF SOLVENT-WATER SEPARATION

None of the methods outlined above solves the problem of how to eliminate water from sources other than stripping steam. Since control of water concentration must be provided for in any case, water originating from the use of stripping steam merely increases the size of the equipment required for the separation of solvent and water. The many advantages of using stripping steam are such that the added cost of equipment mentioned is relatively unimportant.

CLASSIFICATION OF SOLVENTS

For convenience of discussion, the solvents under consideration have been grouped according to the following classifications:

1. Solvents which do not form constant boiling mixtures with water.
   a. Solvents immiscible with water.
   b. Solvents completely miscible with water.
2. Solvents which form a constant boiling mixture with water where the mutual solubility of the two is small.
3. Solvents which form a constant boiling mixture with water where the mutual solubility of the two is appreciable.

The data given in Table 1 forms the basis for this classification. When mixtures of solvents are used, the solution of the problem of separation will depend on which component of the solvent mixture is associated with water.

TABLE 1

Properties of Commercial Solvents

Solvent	Boiling Point	Solubilities at 100° F.		Constant Boiling Mixture, Boiling Point	Composition Constant Boiling Mixture Overhead	Classification
		Solvent in Water	Water in Solvent			
	Deg. F.	Wt. %	Wt. %	Deg. F.	Wt. % Solvent	
Acetone	133	0.0	0.0	none	none	1b
Aniline	364	3.8	5.6	209	23.1	3
Benzol	176	0.16	0.11	none	none	1a
Carbon textrachloride	170	0.09	0.01	none	none	1a
Chlorex	353	1.2	1.1	204–207	.....	2
o—Cresol	375	3.0	14.5	210	6	3
m—Cresol	396	2.5	14.5	210	6	3
p—Cresol	396	2.2	16.5	210	6	3
Ethylene dichloride	182	0.9	.....	158	91.8	2
Furfural	323	9.0	6.5	208	35	3
Methyl ethyl ketone	175	19.0	10.2	164	88.6	3
Nitrobenzene	412	0.3	0.35	211	15.3 (14)	2
Phenol	358	9.4	32.5	210	9	3
Propane	—48	3.4	.....	none	none	1a
Sulphur dioxide	14	...	.....	none	none	1b
Toluol	231	0.05	.....	none	none	1a
Trichloroethylene	188	0.16	0.05	167		2

Separation of Water and Solvents of Classification 1.

The solvents listed under Classification 1-a are benzol, toluol, liquid propane and carbon tetrachloride. Due to the extremely low mutual solubility of these solvents and water, separation by decantation is entirely satisfactory and losses of solvent in water are negligible.

The solvents listed under Classification 1-b are acetone and liquid sulfur dioxide. Since acetone is miscible with water in all proportions, the separation is effected by simple fractionation in a suitably designed fractionating column. Due to the chemical reactions involved, the classification of liquid sulfur dioxide given is not entirely correct, but will serve for the present discussion. The characteristics of liquid sulfur dioxide are such that special precautions must be taken to avoid the presence of water in the system. The necessity for these precautions is evident when consideration is given to corrosion difficulties which would otherwise result. Normally, all entering streams are carefully dehydrated and water cooling equipment is constructed to avoid any possibility of leakage.

One well known commercial dewaxing process uses a solvent mixture containing acetone and benzol. This mixture offers an interesting problem of water separation since it is possible to have a mixture of the two and water. In the absence of acetone, separation of the mixture results in a benzol phase and a water phase. However, acetone is soluble in both benzol and water, and by distributing itself between the two phases increases their mutual solubility. To avoid complications in recovery equipment resulting from this, the concentration of acetone in the water phase is kept below the point where any appreciable solubility of benzol in the water phase would result. The special problem then ceases to exist.

Separation of Water and Solvents of Classification 2.

Solvents found under Classification 2 are Chlorex, ethylene dichloride, nitrobenzene, and trichloroethylene. Since most of the solvent is recovered free from water, the small proportion associated with water is passed to a separator. The solvent phase saturated with water is added directly to dry solvent since the water it contains is insufficient to affect the water content of the total solvent to a significant extent. The water phase saturated with solvent is processed as shown in Figure 4.

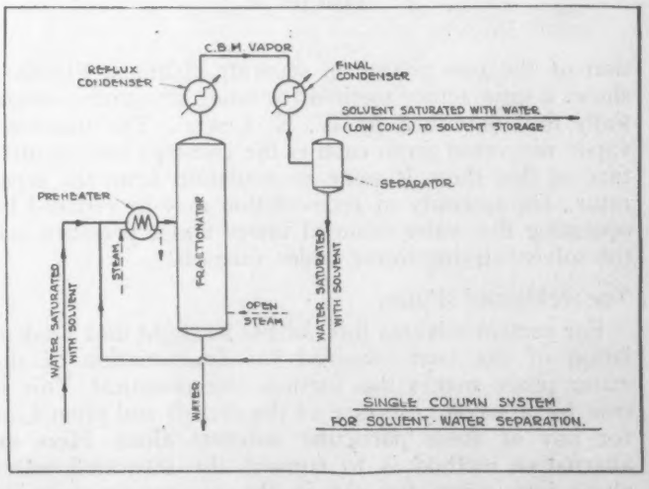


FIGURE 4

The saturated water is preheated and discharged into the fractionating column, from which the constant boiling mixture is removed as overhead vapor and water free from solvent as residue. Steam is admitted directly into the bottom of the column to supply necessary heat.

Separation of Water and Solvents of Classification 3.

Most of the commercial solvents occur in classification 3. Among these are aniline, o-cresol, m-cresol, p-cresol, furfural, methyl ethyl ketone and phenol. The methods used for separating water and solvents under this classification are varied and the problem has many interesting solutions.

Furfural and Water.

Considering furfural and water we find that the constant boiling mixture consists of about 35 percent furfural and 65 percent water. Condensation and cooling of this mixture yields two phases. The upper phase consists of water containing about nine percent furfural while the lower phase consists of furfural containing about 6.5 percent water. One factor of importance is that by concentrating furfural into a constant boiling mixture the quantity of material to be handled is a minimum.

For the example chosen, there is a considerable shift in the solvent concentration between that in the constant boiling mixture and that in the solvent phase of the cooled condensate. This shift is sufficient to enable recovery of each component as a residue by fractionation of the two phases in separate columns. Figure 5 shows a satisfactory method for doing this, and is essentially that described by W. K. Lewis[7]. The overhead vapor recovered from each is the constant boiling mixture so that there is some recirculation from the separator. The quantity of recirculation may be reduced by operating the water-removal tower under pressure and the solvent-drying tower under vacuum.

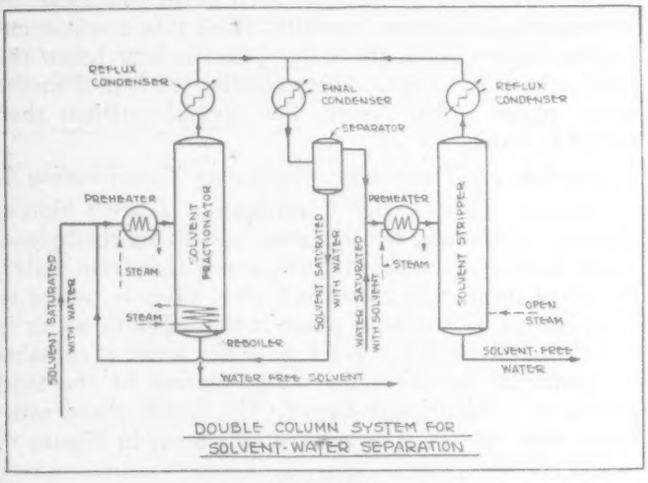


FIGURE 5

Tar Acids and Water.

For certain solvents this shift is so slight that a calculation of the heat required for fractionation of the water phase makes this method uneconomical. This is true for a solvent mixture of the cresols and phenol, or for any of these particular solvents alone. Here an alternative method is to convert the saturated water phase into steam for use in the process as stripping steam[8]. This has been proved to be entirely practical in commercial application because of the difference in molecular weights of the solvent and water and the fact that the stripping operation is carried out under vacuum at an elevated temperature. To take actual figures, let us assume the stripping operation is carried out at 100 mm Hg. absolute pressure. The water phase used for stripping steam contains 2.5 percent by volume of solvent and 97.5 percent by volume of water. Inasmuch as the molecular weights of the solvent and water are in the ratio of six to one, we find that the mol percent of solvent in the vapor mixture amounts to 0.42 percent. Used as stripping steam at the absolute pressure indicated, the partial pressure of solvent is only about 0.4 mm. At the temperature employed in the stripping operation the reduction in efficiency of stripping resulting from this small partial pressure of solvent is negligible. It should be noted that although this method takes care of the water present in the system by reason of steam stripping, it does not eliminate possible losses caused by water entering the system from an external source.

Salting Out Method.

An extraction method used for many years in the chemical industry is based on the principle of "salting out." Various salts may be used, sodium sulphate, sodium chloride, etc. Selection of the most satisfactory salt can be made by determining the solubility of the solvent in salt solutions of varying concentrations. The operation may be carried out either as a batch or continuous process. The exact method used probably will depend on economic factors. One continuous system is illustrated in Figure 6. Referring to the diagram, the water phase is mixed with a solution of sodium chloride. The concentration of salt is maintained slightly below the saturation point to avoid deposition of salt in the equipment. By controlling the proportions of salt solution and water phase the dilution caused by the introduction of the water will not be great enough to allow any appreciable quantity of solvent to dissolve in the diluted salt solution. After providing sufficient contact to reach substantial equilibrium, the mixture is fed into a settling tank and allowed to separate. The upper phase consists of the solvent containing about 0.5 percent water, while the lower phase consists of salt solution with a trace of solvent. If much salt is used, the diluted salt solution can be re-concentrated to its original strength and circulated in a closed system. The losses from this operation are small and the operation itself

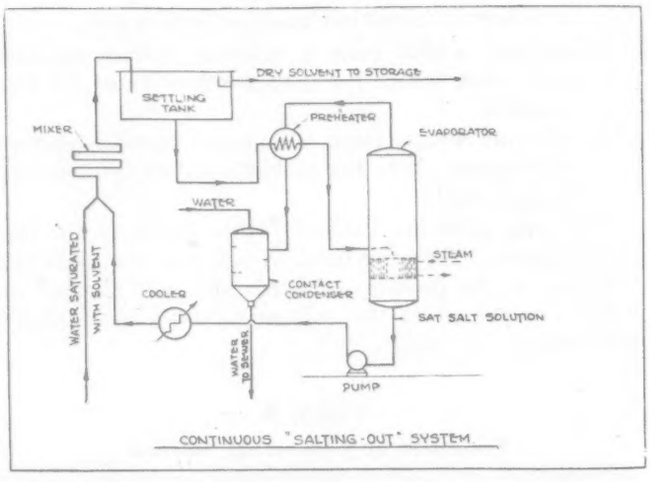


FIGURE 6

easily controlled. Necessary heating can be obtained by the use of exhaust steam because of the low temperature level.

METHOD ADDING LIGHT DISTILLATE

A distillation method which avoids the formation of solvent-water mixtures has been tried experimentally and found to be very practical. This method is

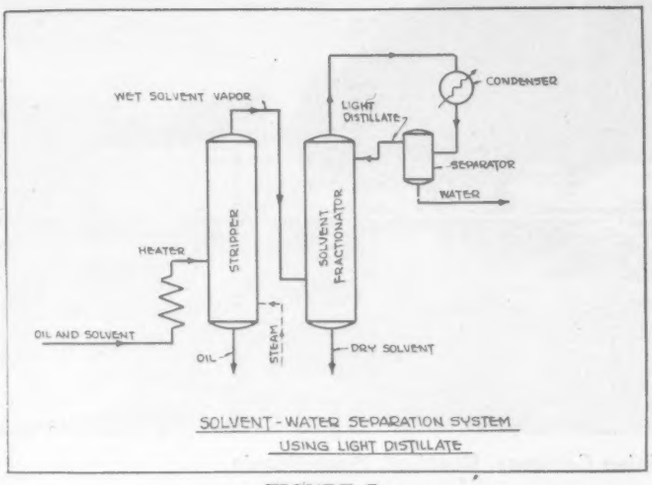


FIGURE 7

shown in Figure 7. The overhead vapor from a stripping tower such as shown in Figure 1 passes to a second fractionating tower. Fractionation of this mixture ordinarily would result in dry solvent as a residue and the constant boiling mixture of solvent and water as an overhead product. By adding a suitable third component with a boiling point below that of the constant boiling mixture, the overhead vapor will consist of this component and steam. Addition of the third component can be taken care of by simply supplying it as reflux at the top of the column. A satisfactory material for use in this method would be a light hydrocarbon of comparatively narrow boiling range. After condensation and cooling of the overhead vapor, the hydrocarbon is returned to the top of the column as reflux while the condensed steam is withdrawn and discarded. It can be seen that the amount of this added material required is small and the possible losses negligible.

MISCELLANEOUS METHODS

Methyl ethyl ketone, a dewaxing solvent, presents difficulties typical of these separating problems. In this case the quantity of solvent which is associated with water is extremely small (less than 0.5 percent) when compared to the total solvent used in the process. Thus, although the solvent phase which separates on condensation and cooling of the constant boiling mixture contains approximately 10 percent water, it is added directly to dry solvent.

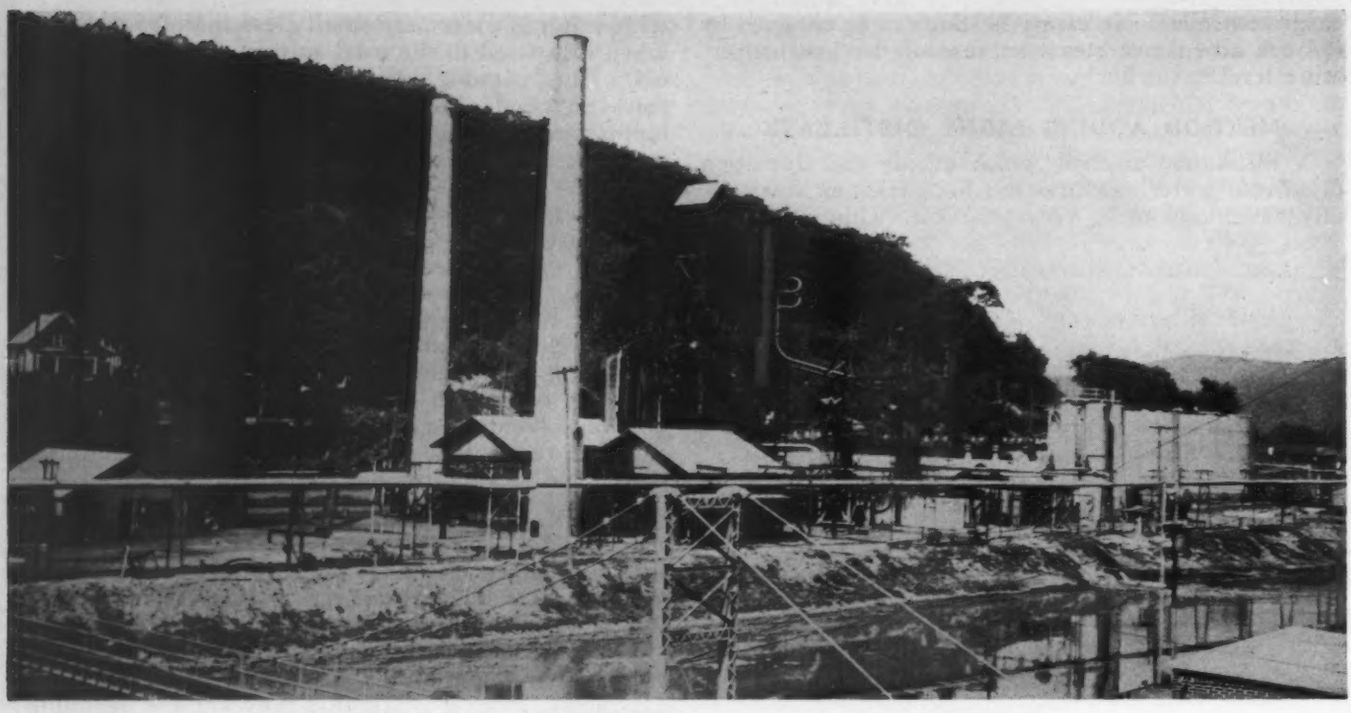
Several methods for removal of solvents from water employ an extraction step in addition to fractional distillation. One such method involves the extraction of water saturated with solvent by a light distillate. The resulting distillate phase contains most of the solvent and can be fractionated to recover light distillate as an overhead product and solvent as a residue. The water phase remaining after the extraction contains a negligible concentration of solvent and is discarded. A similar method employs raw oil charge as the extracting material to remove solvent from water.

SUMMARY AND CONCLUSIONS

In this paper the authors have indicated the problem of water-solvent separation in solvent refining and dewaxing plants and have discussed a number of solutions. Processes involving the use of chemicals or absorbants for water separation have not been considered, since it is felt that they are less desirable in oil refinery operation. The authors have listed certain solvents, analyzed their characteristics, and have shown how the variation in characteristics affects the choice of method to be employed. In general, fractional distillation appears to be most satisfactory because of its adaptability to solvent recovery systems and because of the familiarity of the refiner with this type of equipment. In most cases the heat required may be provided economically with available exhaust steam. In conclusion it may be said that the choice of method used depends on the individual problem, with the decision resting on a consideration of all factors involved.

BIBLIOGRAPHY

[1] Bahlke, Brown, and Diwoky, Proc. Am. Pet. Inst., 14 (III), 77, (1933.).
[2] Bahlke, Giles, and Adams, ibid, 14M (III), 16, (1933).
[3] Bray, Swift and Carr, ibid, 14 (III), 96, 1933).
[4] Bryant, Manley, and McCarty, ibid, 16M (III), 127, (1935).
[5] Ferris, Myers, and Peterkin, ibid, 14 (III), 63, (1933).
[6] Govers and Bryant, ibid, 14M (III), 7, (1933).
[7] Lewis, W. K., U. S. Pat. No. 1,911,832.
[8] Lummus, W. E., U. S. Pat. No. 1,603,851.
[9] Poole and Wadsworth, Proc. Am. Pet. Inst., 14 (III), 39, (1933).
[10] Stines, D. E., U. S. Pat. No. 2,006,186.
[11] Stratford, Pokorny, and Huggett, Proc. Am. Pet. Inst., 14 (III), 90, (1933).
[12] Tuttle, ibid, 16M (III), 112, (1935).
[13] Tuttle and Miller, ibid, 14 (III), 85, (1933).
[14] Wells, Rhodes, and Murray, Ind. Eng. Chem. 17, 1199, (1925).
[15] Williams, D. B., Proc. Am. Pet. Inst., 16M (III), 104, (1935).
[16] Wilson and Keith, ibid, 15 (III), 106, (1934).
[17] Ziegenhain, Oil and Gas J., 28 30-125, (1929).
[18] Pester, Proc. Am. Pet. Inst., 14M (111), 4, (1933).
[19] Cottrell, ibid., 14 (III), 60, (1933).



Cracking plant in refinery of Kendall Refining Company, Bradford, Pennsylvania

Coking Operation

168 Days on Stream

J. C. ALBRIGHT

KENDALL Refining Company, Bradford, Pennsylvania, was one of the first Pennsylvania refiners to install facilities for cracking gas oil, fuel oil, and other heavy by-products, into premium grade motor fuel. The first installation of this character consisted of two Dubbs units, erected in 1923, and the third unit was placed in operation during 1931, which was designated as Plant Number Three. It was operated in conventional manner, with minor changes now and then, until the first part of 1934, when changes completed converted it from a residuum, or coking plant, to one which produces coke outside the present equipment. The results of these changes, demonstrated by a record run recently completed, was to enable the plant to remain on stream for a period of 4046 hours and 45 minutes without a shut-down.

Two unused towers previously had been erected in the cracking plant which contained clay for treating the pressure distillate in the vapor stage. Because of other improvements these became inactive and were considered obsolete material. These two clay towers were converted to chambers for outside coking during 1934. The first of the columns was converted to its present use to demonstrate the feasibility of that type of operation, and after satisfactory experimental runs, the other chamber also was converted.

After all connections and other fittings had been made to convert the cracking plant so outside coking could be accomplished with two chambers hooked up for operation, a run was begun to determine how long the plant could be operated. This run continued for a period of 2,360.65 hours, with several stocks being charged to the coils in the heaters, the results of which was that 133,646 barrels were processed with a recovery of 74.8 percent gasoline, having an octane number of 64.5, C.F.R.

A summary of this run is shown in Table 1.

The plant was taken off stream because of a leak developing in a tube in the heavy oil furnace where it had been rolled into a header. The total coke production during this run amounted to 1949 tons. The run also demonstrated that with the operation of the two outside coking chambers, it was possible to increase the gasoline production as much as six percent, when compared to the method of operating the plant before it was changed over to outside coking.

Of the total time on stream for this outside coking run, 23 hours were consumed in starting up the unit and in bringing it off stream. During this time, the coking operation was suspended, and the bottoms run to residuum. The coke deposition in the furnace tubes varied from dust in places, to not to exceed a thickness of 1/64-inch, and approximately 500 pounds were distributed over the wall surfaces of the various parts of the plant. In the reaction chamber there was no more coke deposited on the inside walls, apparently, than

TABLE 1
Yields and Throughput for Complete Run

	Barrels	Percent of Raw Oil	Barrels per Day
Light charge	23,956	....	244
Heavy charge	109,690	....	1,115
Total charge	133,646	....	1,359
Gasoline produced	99,934	74.8	1,016
Residuum	346	0.3	4
Coke	11,044	8.2	112
Gas and loss	22,324	16.7	227

Gas produced	114,885,900 cu. ft.
Cubic feet of gas per barrel of charge	859
Coke per barrel of charge	30 pounds

Distillations

	Heavy Feed	Light Feed	Side Cut	Coke Dist.	Comb. Feed	Finished Stabil. Gasoline
Gravity	38.6	46.2	19.3	16.6	14.7	61.0
I. B. P.	370	338	260	.....	.....	83
10 percent	519	394	442	.....	.....	110
50 percent	590	423	504	.....	.....	247
90 percent	690	466	578	.....	.....	379
E. P.	734	510	638	.....	.....	411
Percent received	97	98	99	.....	.....	97
Percent at 400° F.	0–5	16	.....	.....	.....	.....
Percent at 572° F.	37.0	.....	.....	.....	.....	.....
Carbon res.	0.058	.....	0.007	0.073	0.339	.....
Pour	40	.....	.....	.....	.....	.....
B. S. & W.	Trace	.....	.....	.....	.....	.....

Color	15
Gum	3 mgs.
Octane, C. F. R.	65.5
Vapor pressure	10.6 Reid
Time on stream	2,360.65 hours
Expressed in days	98.36

when the plant was first placed on stream at the beginning of the run. This amounted to about one-inch at that time.

At the time the cracking plant was constructed, only one furnace was built for heavy oil service, which is a radiant heating type, equipped with 86 four-inch outside diameter tubes, with an effective operating length of 30 feet. These tubes have a wall thickness of 3/8-inch, and 18 of them were installed in the convection bank of the furnace, 30 in the roof, and 17 in the back wall. A floor tube installation included 21 tubes of the same size as the others.

A second furnace was completed in 1933, also designed by Alcorn Combustion Company, being identified as a double end-fired unit having 16 corborundum tunnels, eight on each side. At this time, several other changes were made in the plant, one in particular being a connection made in the fractionating column, on the side, so a stream of light oil could be removed to be used in the second furnace with other products, for selective cracking. This furnace was equipped with 47 tubes, having an outside diameter of three inches, and an inside diameter of 2¼-inches, located in the convection section. Thirty-three tubes of the same size and wall thickness were placed in the heating section of the furnace, across the floor and in the roof in single rows. Fifty-eight tubes were placed in the soaking side of the furnace, located across the floor and across the roof in double rows, staggered from upper to lower row in both the roof section and the section in the floor. These tubes are slightly larger than the others in this furnace, being 3½-inches, outside diameter, and 2¾-inches, inside diameter. All tubes in this furnace have an effective heating length of 22 feet.

The two clay towers, now termed coking chambers, were prepared for service by building a manhole in the

Interior of control room

lower head, 48 inches in diameter, for ease in removing the coke, and attaching lugs to the interior of the walls so the coke cable could be hung on wires. The necessary piping was done to enable the residuum from the cracking operation to flow to these chambers, manifolded so each could be operated independently. A vapor line was led from these coke chambers to a coke distillate knockout tower, four feet, four inches in diameter by 17 feet in height, equipped with four sets of side-to-side baffles.

After the record run of 2,360.65 hours had been completed, the plant cleaned up and inspected, it was placed on stream again at 9:30 a.m. May 13, 1935, to be operated for coke manufacture by using the coking chambers and to secure the greatest possible yield of cracked gasoline. The fires were cut from the two furnaces at 12:15 a.m., October 29, 1935, after being on stream 4046 hours and 45 minutes. This time, reduced to days, means that the plant operated for 168.61 days, or for five months and 18 days.

Summary of the operation, and the composition of charging stocks is shown in Table 2.

The flow of oil through this cracking unit begins when the charging pump picks up the blended charge of gas oil, petrolatum or foots oil and pumps it through the absorption, or vapor recovery plant column where the gases removed from the accumulator drum and coke distillate receiver, are stripped of the desirable fractions. The charge is handled by a steam driven pump, 7½x4½x10-inch, which delivers the raw oil charge to the cracking unit fractionating column and from this point, oil is taken through a suction line leading to a cross compound hot-oil pump, 23x32x4½x21-inch which delivers the oil directly to the tubes contained in the heavy oil furnace, with the oil being under a coil inlet temperature of about 600°F.

The heavy oil charge being cracked under the in-

TABLE 2
Composition of Charge

	Barrels	Percent
Waterwhite distillate	48,144	18.7
Gas oil	183,296	71.4
Petrolatum	16,478	6.4
Total	256,498	100.0

Trap and sump oils were fed back as recycle stock with the charge, and which amounted to 3296 barrels. The average rate of charge, expressed in fresh oil barrels to the plant amounted to 1521 barrels daily.

Tests on Oil Charged to the Unit

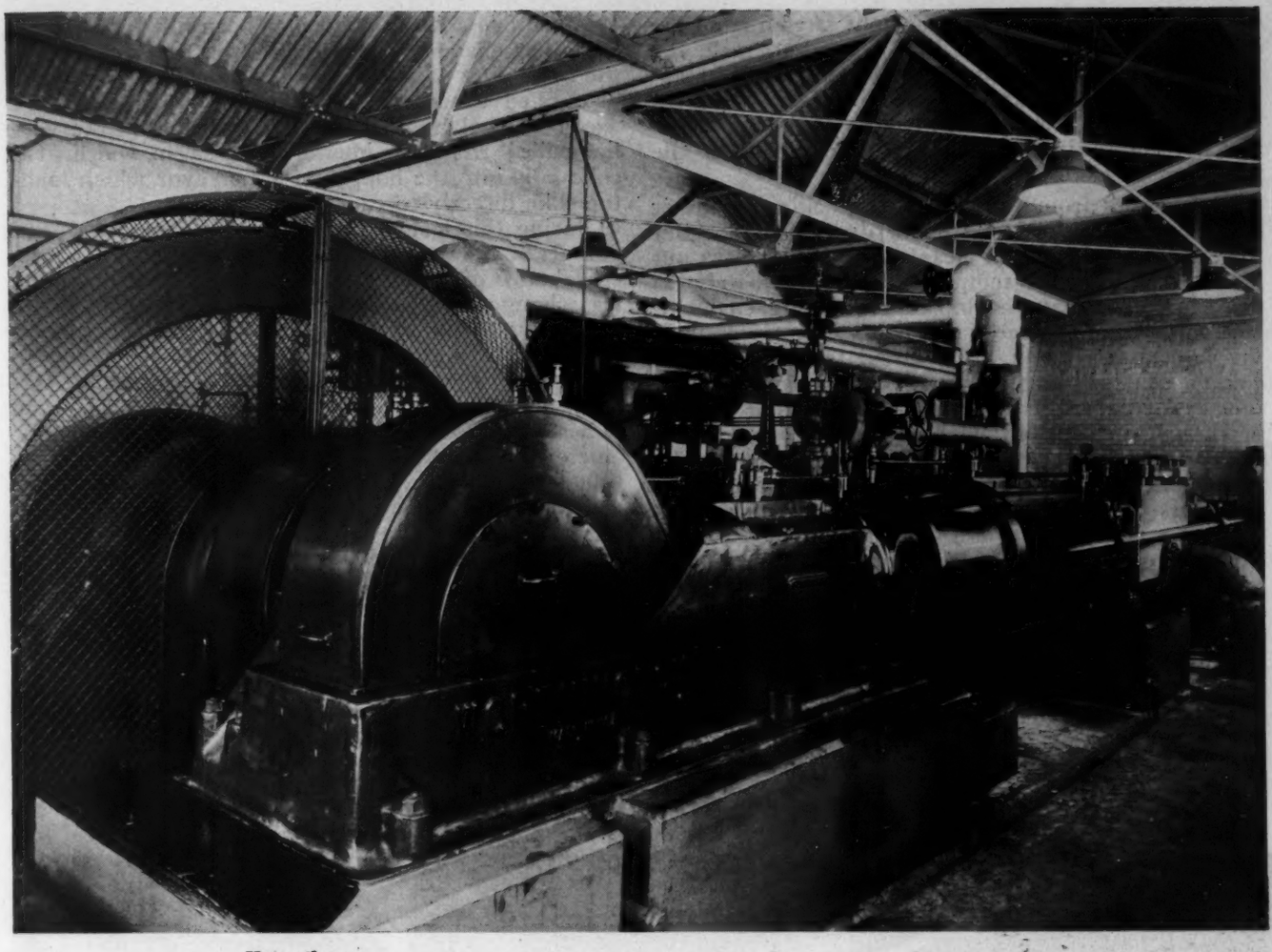
	Gr.	IBP.	10 Percent	50 Percent	90 Percent	E.P. ° F.	C.R.	C.T. and Amp.
Gas oil and Foots oil, mixed before charging	38.6	340	496	586	684	728	0.034	30
Petrolatum (10 mm. hg.)	33.0	564	603	724	....	....	1.100	126
Waterwhite distillate	46.1	380	408	427	460	507	....	....

Yields

	Barrels	Percent
Gasoline (66 octane, 416° F. E. P.)	189,064	73.7
Residuum (2.0 gr. 40 vis. at 122° F. (Furol)	1,913	0.7
Coke	20,683	8.1
Gas and loss	44,838	17.5

Gas produced	199,138,800 cu. ft.
Per barrel charge (gas production)	775 cu. ft.
Fuel consumption	9.9 percent of raw oil charged
Reflux ratio	3.7 to 1

Control board



Hot oil pump

fluence of the furnace temperature and pressure is led through a transfer line at a temperature of approximately 900°F.; all temperatures varying with the type of cracking and the characteristics of the charge used in the plant. The material leaving the transfer line is delivered to the top of the reaction chamber at a pressure of 400 pounds, gauge.

A side stream of lighter oils than the raw charge to the plant, is taken from the side of the fractionating column and picked up by a duplex plunger pump, 3½x12-inch, driven through a set of speed-reduction gears by a vertical single-cylinder steam engine.

The side cut taken from the fractionating unit is passed through a heat exchanger, and may be mixed with any gasoline, naphtha or kerosene that is to be cracked, or reformed, the cut temperature on the inlet of the exchanger being in the neighborhood of 480°F., and passing out at approximately 680°F. From the exchanger, the heated charge enters the convection bank of tubes in the light oil furnace, the flow following then through the floor tubes and the roof tubes in the heating section. The temperature is varied in the heating section to suit conditions found in the plant or charging stock, or the characteristics of the products desired from cracking. The flow continues from the heating section through the soaking tubes in the roof and then through the same section in the floor.

The temperature of the material in the transfer line from the light oil cracking furnace may run as high as 1025°F. at times, and the pressure carried to 700 pounds gauge, depending upon the stock charged to the furnace coils on the light oil side of the cracking system. The products of cracking in both the heavy and light oil furnaces are compined in a common stream, and pass downward through the reaction chamber at a temperature of about 900°F., and under a pressure of 400 pounds, gauge.

The charge passes out the reaction chamber at the bottom through a remote-controlled valve into the lower part of the flash chamber. The residue from the flash chamber is carried to the coking chambers, one of which is always on stream, except when residuum operation is being practiced.

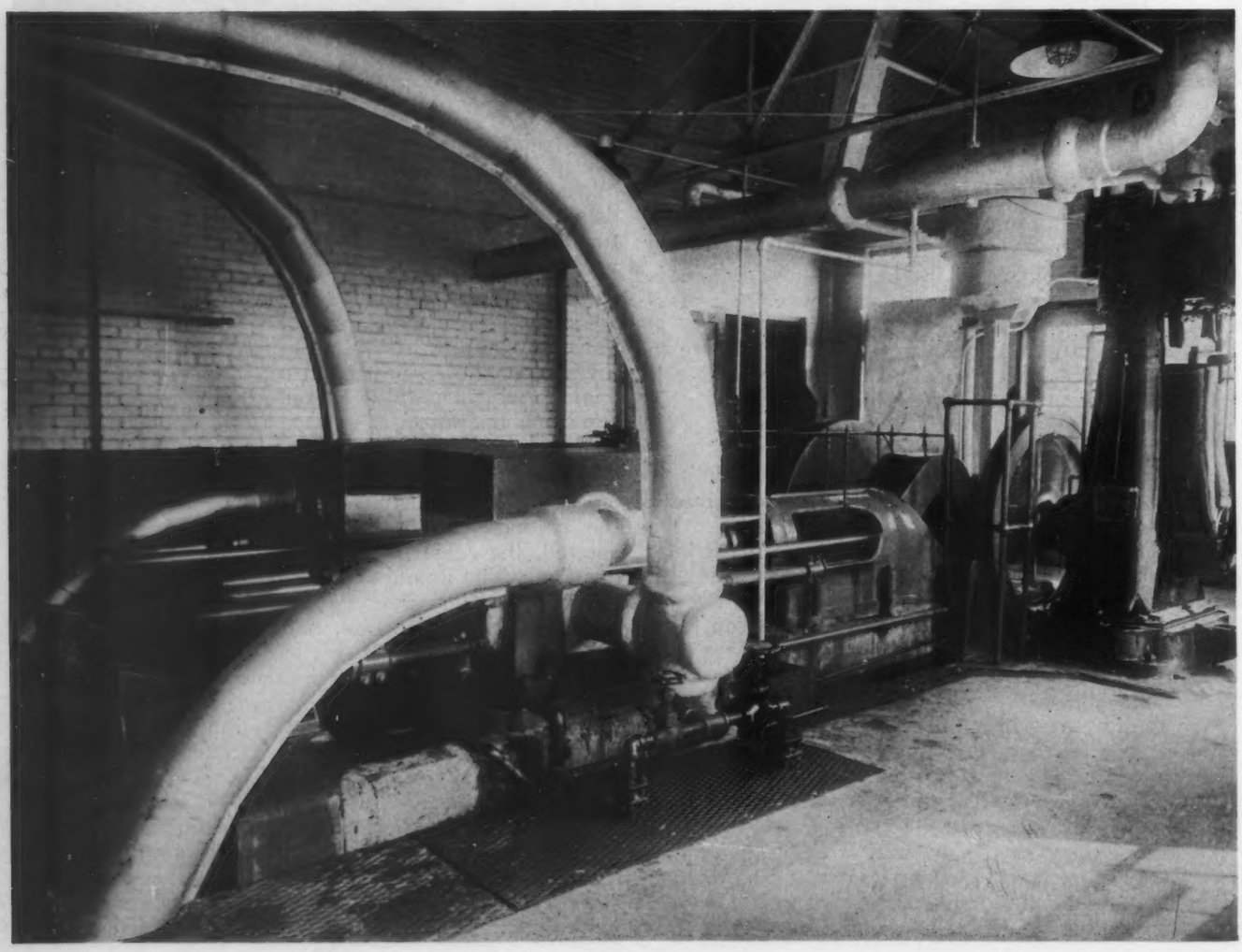
As soon as the cracking unit is brought up to the desired temperatures and pressures and is on stream, operating in the usual manner for residuum, one of the coking chambers, previously having been inspected, the cable hung on the lugs, is steamed to expel air. The amount of cable used in each of these coking chambers averages approximately 8500 feet, and is ⅞-inch wire line made expressly for coking chamber operation. After the air has been expelled, a small amount of the hot material from the transfer line of the light oil cracking furnace is introduced into the bottom of the coking chamber to dry up any water which may have been condensed from the steam used in expelling the air. As soon as the chamber is deemed sufficiently dry, the hot residuum produced from the cracking of the light and heavy charges is switched directly from the bottom of the flash column to the coking chamber and a small stream of hot material from the transfer line on the light oil cracking furnace is constantly fed to the

chamber at the same time. The split stream from the light oil cracking transfer line is used to raise the temperature of the residuum and maintain it at approximately 820°F., and the chamber is carried at a pressure of about 50 pounds, gauge. The addition of the hot light oil stream is sufficient to flash the residuum to coke.

The vapors from the top of the coking chamber are passed directly to a coke distillate knockout tower, where they enter near the base and pass out at the top under a temperature approximately 75 degrees cooler than the top of the coking chamber. Bottoms and other material which will not pass from the knockout column are recirculated back to the coking chamber together with the residuum from the flash column. The vapors removed from the top of the coke-distillate knockout column are passed first through a heat exchanger counter current to the raw charge to the unit on the heavy oil side, and from this exchanger, the vapors and condensate pass through a parallel condenser coil and into the coke distillate receiver. The coke distillate is pumped directly from the receiver drum to the plant fractionating column, and the heavy fractions not carried over as cracked gasoline are passed back as recycle stock for further cracking. The gases which do not condense in the coke distillate cooling sections pass with other vapors to the absorption column in the recovery plant.

When one coke chamber has been filled with coke, the other chamber is treated as the first one was at the beginning of the run and then placed in service. The filled chamber is handled in much the same way other coke chambers are handled, and the wire line pulled from the interior by a steam winch. The coke falls into the hopper of a conveyor which loads the coke directly to cars on a siding.

During this last record run of the plant, 7,300,000 pounds of coke was produced, (equivalent to 42 coke chambers) of good salable quality. The operation was changed from coking to residue and back again to coking seven times during this run for various reasons. The run was terminated only when a tube ruptured in the radiant section of the heavy oil furnace, and which had been in service for four years. When calipered in the usual manner an internal diameter which would still be considered safe for continued operation, was indicated as there was no coke present in this tube nor in any of the tubes in this furnace except for a 1/64-inch scale in the last 10 or 12 tubes in the bank.

A condition had developed in this section of the furnace in which the tubes were worn, or had eroded eccentric, instead of concentric. This condition had not been previously experienced in this unit, therefore, this tube rupture could have happened on a run of much shorter duration and would not have necessarily been anticipated. The condition of the unit otherwise indicated that it could have been carried on stream indefinitely had it not been for this mechanical difficulty.

Light oil cracking charge pump, steam driven through reduction gear